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Development, validation, and use of a spreadsheet-based tool for early-stage technoeconomic evaluation of industrial biotechnologies

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**Development, validation, and use of a spreadsheet-based tool for early-stage
technoeconomic evaluation of industrial biotechnologies**

by

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A thesis submitted to the graduate faculty

in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

Major: Agricultural Engineering

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TABLE OF CONTENTS

List of Figures	iv
List of Tables	vii
Abstract	ix
Chapter 1: Introduction and Literature Review	1
Objectives.....	6
Thesis Organization.....	7
Author's Role	7
References	8
Chapter 2: A Coarse Technoeconomic Model of a Combined Fermentation-Catalysis Route to Sorbic Acid	11
Introduction	11
Materials and Methods	13
Processes	14
Model	16
Results	27
Discussion	35
Conclusions	36
References	38
Chapter 3: Development of a Biorenewables Process Evaluation Tool: BIOPET	41
Introduction	41
Methods.....	43
Model Comparison: Approach.....	56
Model Comparison: Ethanol	57
Model Comparison: Succinic Acid	59
Model Comparison: Adipic Acid	61

Ethanol Comparison.....	62
Succinic Acid Comparison.....	64
Adipic Acid Comparison.....	65
Discussion	67
Conclusion.....	72
References	73
Chapter 4: Technoeconomic evalution of Bio-Based Styrene production from <i>Escherichia coli</i>	
Introduction	77
Methods.....	80
Results and Discussion.....	82
Conclusion.....	87
References	88
Chapter 5: General Conclusions	91
References.....	93
Acknowledgements.....	101

LIST OF FIGURES

Figure 1: Simplified process diagram for proposed sorbic acid process and associated mass flows. Red lettering indicates mass flows leaving the system.	16
Figure 2: Summary of scenarios investigated in this work showing key parameter values. Changes between scenarios are highlighted in yellow text.	26
Figure 3: Cost Distributions for Three Scenarios in the Current Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.	27
Figure 4: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the current case of sorbic acid	28
Figure 5: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the current case of sorbic acid	29
Figure 6: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the current case of sorbic acid	29
Figure 7: Cost Distributions for Three Scenarios in the Short-Term Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.	30
Figure 8: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the short-term case of sorbic acid	30
Figure 9: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the short-term case of sorbic acid	31
Figure 10: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the short-term case of sorbic acid	32

Figure 11: Cost Distributions for Three Scenarios in the Long-Term Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.	33
Figure 12: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the long-term case of sorbic acid	33
Figure 13: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the long-term case of sorbic acid.....	34
Figure 14: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the long-term case of sorbic acid.....	34
Figure 1: CBiRC's focus in chemical commodity value chain.....	43
Figure 2: SuperPro process flow diagram for ethanol	58
Figure 3: SuperPro process flow diagram for succinic acid	60
Figure 4: SuperPro process flow diagram for adipic acid.....	61
Figure 5: Comparison of the economics between BioPET and SuperPro process models for ethanol.....	64
Figure 6: Comparison of the economics between BioPET and SuperPro process models for succinic acid.....	65
Figure 7: Comparison of the economics between BioPET and SuperPro process models for adipic acid	67
Figure 9: Regression of predicted selling price for ethanol, succinic acid, and adipic acid for BioPET and SuperPro Designer® modeling tools (SuperPro error bars, $\pm 25\%$; BioPET error bars, $\pm 35\%$).....	67

Figure 10: Sensitivity of production capacity to selling price for ethanol, succinic acid, and adipic acid using the BioPET tool	69
Figure 11: Sensitivity of feedstock cost for ethanol, succinic acid, and adipic acid using the BioPET tool	70
Figure 12: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for ethanol.....	71
Figure 13: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for succinic acid.....	71
Figure 14: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for adipic acid.	72
Figure 1: Top five sensitivity coefficients using a $\pm 1\%$ change in BioPET model under expected values for bioproduction of styrene (x-axis represents measured change in MESP). ...	85

LIST OF TABLES

Table 1: Key Economic and Plant Performance Parameters	19
Table 2: Fermentation Parameters	21
Table 3: Separation Parameters	23
Table 4: Catalysis-Hydrogenation Parameters.....	24
Table 5: Catalysis - Solid Acid Parameters	24
Table 6: Purification Parameters.....	25
Table 7: Prices of key process inputs.....	25
Table 1: Separation and Purification Unit Operations.....	44
Table 2: Input Variables Necessary for Unit Operations.....	44
Table 3: Fermenter Assumptions	46
Table 4: Centrifuge Assumptions	48
Table 5: Adsorption Assumptions	49
Table 6: Crystallization Assumptions.....	51
Table 7: Distillation Assumptions	52
Table 8: Liquid-Liquid Extraction Assumptions	53
Table 9: Catalysis Assumptions.....	55
Table 10: Cost of Key Materials and Supplies	56
Table 11: BioPET Inputs – Ethanol	58
Table 12: BioPET Inputs – Succinic Acid.....	60
Table 13: BioPET inputs – Adipic Acid.....	62
Table 14: Comparison of SuperPro and BioPET output for ethanol production.....	63
Table 15: Ethanol Results: BioPET	63

Table 16: Succinic Acid.....	64
Table 17: Succinic Acid Results: BioPET	65
Table 18: Adipic Acid.....	66
Table 19: Adipic Acid Results: BioPET	66
Table 1: Base-Case (expected value) and range of parameter values for Monte-Carlo analysis of the economics of styrene bioproduction.	82
Table 2: Capital Expenditures.....	83
Table 3: Annual Expenses	84

ABSTRACT

This thesis begins with a technoeconomic analysis (TEA) of bio-based sorbic acid (SA). The initial TEA focused only on SA, and did not attempt generalizations. However the experience of building the SA TEA inspired development of a more general tool designed for early-stage TEA's of hybrid biological/chemical systems for producing bio-based chemicals, as proposed by the founders of the NSF Engineering Research Center for Biorenewable Chemicals (CBiRC). This early-stage TEA tool, named BioPET (Biorenewables Process Evaluation Tool), was designed for the purpose of conducting rapid early-stage economic analyses of these hybrid systems. BioPET was validated against a commercial economic analysis tool, SuperPro Designer[®], and against published literature. BioPET was subsequently used to evaluate a recently developed pathway for bio-based styrene. BioPET fills a critical niche in the evolving bio-based chemicals industry. This is because there is a need for low-cost tools capable of early-stage estimations of the economics for novel bioprocess systems. Such a tool can provide valuable insight into nascent projects.

This thesis is prepared in paper format, and is comprised of three manuscripts, as follow: The first paper was an evaluation of the economics of bio-based sorbic acid production. Sorbic acid has a growing market in food preservatives mainly due to health concerns about benzoic acid, currently-used food and cosmetic preservative. While bio-based sorbic acid has reached proof-of-concept stage, little has been done to understand the costs of a commercial-scale process and the economic feasibility of such a venture. A spreadsheet model was created for the purpose of conducting this evaluation and understanding how critical biokinetic parameters influence the final estimated selling price. Based on current

values of these parameters, we appear far from producing a product that can be sold at commercial scale. However, by assuming improvements in key parameters that reflect experience with other fermentative systems, bio-based sorbic acid becomes cost competitive with current petroleum-based sorbic acid. Production costs were most sensitive to those parameters governing the overall yield of sorbic acid in this process. In the long-term projection, primary costs were almost equally shared amongst feedstock, separation, and catalysis. Improving yields for this process will be required to make this process economically feasible, but also vital will be improving all kinetic parameters in order to achieve cost competitiveness.

The second paper explored the development of a robust but simple spreadsheet model (BioPET) to perform early-stage TEA of candidate processes for biorenewable chemical production. In the early-stage development of new technologies, a feasibility study or order-of-magnitude evaluation TEA is conducted to determine whether further development of that technology is warranted. With the number of new technologies and pathways being developed in the realm of industrial biotechnology, a tool that can provide a rapid estimation of a new technology has great value in delivering feedback to scientists and companies alike. Using basic inputs governing fermentation (e.g. productivity, titer, yield), separation (e.g. distribution coefficient, relative volatility, purity, yield), and catalysis (e.g. selectivity, conversion, type of catalyst), an estimate of a production price can be determined. This early-stage TEA tool was built in Microsoft Excel[®] and evaluated for accuracy and precision against SuperPro Designer[®] and the BREW project from the EU, using ethanol, succinic acid, and adipic acid as target chemicals. Processes were simulated as close to the BREW assumptions as possible. BioPET had accurate results against SuperPro Designer[®], providing

an R^2 between the two tools of 0.9995. BioPET had minor deviances from BREW project projected selling prices of the evaluated chemicals, but the results were within the range of error for BioPET-derived estimates.

The third paper describes the application of BioPET to the evaluation of bio-based styrene. Bio-based styrene is a drop-in replacement chemical that remains in the early stages of development. Given basic knowledge of the properties of chemicals used in the process and general knowledge of the biokinetic limitations of the host organism, the styrene process was evaluated in BioPET at conservative commercial-scale values to evaluate the competitiveness of such a method of production. The results suggest bio-based styrene could be competitive with current petroleum-based prices at predicted selling price of 1.82 USD kg^{-1} . A Monte Carlo analysis provided insight into the uncertainty of the process and estimated an the standard deviation to be ± 0.44 USD kg^{-1} . The majority of the cost of bio-based styrene arises out of the feedstock due to the small maximum yield of fermentation and relatively simple process design. While current production values might not yet be commercially feasible, values of bio-based styrene have potential to surpass the current petroleum-based styrene production. Additional research into the metabolic pathways governing biostyrene production will enable a reduction in the uncertainty of the cost estimate. At present, the BioPET results on bio-based styrene, and rising prices of petroleum-based styrene, suggest that bio-based styrene may well be cost-competitive in the future.

CHAPTER 1: INTRODUCTION AND LITERATURE REVIEW

An increasing population and dependence on products synthesized from a barrel of oil has created a great interest in understanding the total amount of available hydrocarbon resources (Rogner, 1997). It has been suggested that while fossil carbon reserves remain plentiful, the cost of extracting these less readily available resources will be potentially cost prohibitive (Rogner, 2012). This information has led to a surge in research exploring the replacement of the hydrocarbon-derived products with ones derived from renewable carbon (i.e., photosynthetically-derived). With a vast quantity of the barrel of oil producing multiple sources of liquid-fuels, researching bio-based energy has taken precedence over research into other bio-based products (Nikolau et al., 2008). A major portion of the liquid fuels being consumed is utilized in passenger vehicles. One potential replacement for liquid fuels utilizes fermentation of biologically-derived sugars (predominantly from sugarcane in Brazil and from corn-starch in the US) to produce ethanol, a viable fuel for spark-ignition internal combustion engines (Hsieh et al., 2002; Yüksel and Yüksel, 2004). During the US oil crisis in the 1970's, an increased number of investigations were conducted to understand the economics of industrial fermentation (Cysewski and Wilke, 1978; Maiorella et al., 1984). With a subsequent decline in oil prices bringing relatively inexpensive fuels back to the US market, interest in fermentation-derived fuels waned. However, in the early 2000's, a variety of global economic factors and policy responses greatly increased interest in ethanol and in other fermentation-derived compounds. A large number of studies focused on how to extract the sugars for fermentation, and others have focused on process design to improve the economic competitiveness of fermentatively-derived fuel compounds (Huang et al., 2009; Jiang et al., 2011; National Renewable Energy et al., 2000; Qureshi and Blaschek, 2000). Another way of improving the economic competitiveness of

fermentation-derived compounds is to compete with petroleum-based value-added products known as petrochemicals (Landucci et al., 1994; Werpy and Petersen, 2004). While fuel-like compounds have been studied extensively, and have a plethora of process details available in the literature, the process details and economics of value-added products from biomass are not as well understood. This is compounded by the diversity of physical and chemical properties in this realm, making understanding all potential processes is an incremental venture (Choi and Lee, 1997; Datta et al., 1995; Van Wegen et al., 1998).

Production of value-added products from biomass, or bio-based chemicals, has been greatly aided through the advent of a new field known as metabolic engineering. Metabolic engineering employs techniques using recombinant DNA to better understand the organisms utilized for the fermentation processes and aims to improve the economics of such situations by improving stress responses of organisms to their environment, productivities, titers, and yields (Jarboe et al., 2010; Nielsen, 2001; Purvis et al., 2005; Sánchez et al., 2005). Allied fields of biotechnology such as plant biotechnology (sometimes referred to as green biotech) and medical biotechnology (sometimes referred to as red biotech) have made tremendous improvements through similar techniques implemented in their respective field. This field, known as industrial, or white, biotechnology, is making similar improvements but the diversity of compounds and economic challenges of white biotech are arguably greater than in green or red biotech. While red biotechnology may use fermentation, the target compounds are typically extremely high-value, produced in low quantity, and not economically attainable by non-biotech methods (Werner, 2004). In contrast, the products of industrial biotechnology must be economically competitive with their petro-based counterpart. Green biotech is aimed at incremental

improvements to a previous system and therefore does not have to design from the ground up as does industrial biotechnology.

Producing compounds of interest directly via fermentation has been the primary focus of current research in industrial biotechnology, although alternative methods do exist (Choi and Lee, 1997; Kazi et al., 2011; Rivas et al., 2004; Song and Lee, 2006; Straathof et al., 2005). However, fermentative methods for producing bio-based chemicals are capital intensive and financially risky; this has led to the idea of biorefineries. A biorefinery is functionally similar to a petroleum refinery, in that a portion of the incoming feedstock is diverted to fuels, while the other can be diverted to bio-based chemicals (Kamm and Kamm, 2007). This should reduce the amount of capital spent per product developed and decrease risk by having a more diverse portfolio of chemicals. One method of accomplishing this concept of a biorefinery is to develop “platform” chemicals; a chemical that serves as an intermediate to a variety of other chemicals (Nikolau et al., 2008). Work has been done on developing chemicals from both the fermentation and catalytic side that would serve as platform chemicals (Chia et al., 2012; Marr and Liu, 2011; Nikolau et al., 2008)

As the opportunity for chemicals to be produced via fermentation and catalytically manipulated continues to expand, understanding the economics behind the development of such processes could provide insight into potential bottlenecks and limitations of such an approach. Current methods for conducting technoeconomic analyses (TEA), such as Aspen™ and SuperPro Designer®, are extremely powerful, and have the ability to consider detailed mass and energy balances on each unit operation in the modeled system. However, these detailed methods are also relatively time-consuming and limited in scope because of the extensive amount of detail they both require and produce. When limited amounts of data are available, a simple flowchart may

be the best way of capturing uncertainty while still providing an estimate until adequate measures are acquired and can be used in the prior methods (Bunger, 2012). The NSF Engineering Research Center for Biorenewable Chemicals (CBiRC) is working on developing a greater understanding of not just the science between producing these new and novel chemical pathways, but in understanding the economics of producing chemicals through such pathways.

CBiRC is positioned as a research center developing translational technology that aims to transform the chemical industry through the development of platform chemicals. To achieve this goal, CBiRC has created testbeds as proof-of-concept platforms that integrate fermentation and catalytic routes, and as a method of exploring this new research space. One such testbed aims to produce sorbic acid via a combined fermentative-catalytic route. This route aims to examine triacetic acid lactone (TAL) as a potential platform chemical and explore the diverse catalytic opportunities. TAL is within a family of molecules known as pyrone, and a significant amount of work within CBiRC has focused on improving fermentative pyrone production via enzymology and metabolic engineering. Sorbic acid was chosen as a final target product because of its *drop-in* capabilities. This can be beneficial because *drop-in* chemicals have identical chemical formulas and structures to the chemical already in place and are of decreased technical risk to develop. In contrast, chemicals known as *functional replacements* are molecules that offer similar properties as the current market chemical but do not have an identical structure. An example of a functional replacement would be NatureWorks™ poly-lactide molecule, which can be made into plastic utensils capable of replacing polystyrene utensils, but slight differences in properties will exist between the polymers because of the different monomer base for the polymers.

As a translational research facility, improving methods for evaluating the economics of new chemicals via this combined fermentative-catalytic route is a necessity. Initial design will attempt to develop a model for evaluating sorbic acid via this combined fermentative-catalytic route. The nature of translational research inherently limits the amount of available data for estimating process economics, and the main platform for evaluating the economics will be based in Microsoft Excel[®] to meet the desired qualities in this evaluation tool. Using Excel will make the evaluation tool useful to an extremely wide customer base because of the plethora of computers running various versions of Microsoft Excel[®]. The ideal model will offer ease of use, transparency, and accuracy. While a programming language known as Visual BASIC for Applications (VBA) does exist within the Microsoft framework, VBA can limit the transparency, and many operations can be executed without the need for VBA. For this reason VBA was intentionally avoided in the model. This decision also resulted in a model that provides instantaneous feedback as inputs are changed in the cell.

A tool that evaluates the CBiRC sorbic acid process would provide insight into the future economics of bio-based chemicals, but the knowledge provided would be limited to that of a single process. Under this presumption, new spreadsheets or process flow diagrams would have to be created for each new process design and operate under their respective guised assumptions (Choi and Lee, 1997; Maiorella et al., 1984). While often times a TEA only examines a single new technology, considering process alternatives within the same framework can provide more insight into the process and economics of the product and feasibility. Due to the length of time required to conduct a TEA, considering process alternatives becomes difficult and cumbersome due to missing details and process know-how. By providing a single tool that can provide economic evaluations for multiple process designs, the model greatly reduces time to evaluation.

This tool could effectively act as a platform for future evaluations and be adapted as necessary using knowledge of the specific process in mind.

A candidate for a tool as aforementioned would be bio-based styrene from *Escherichia coli* because of its early stage in development. Bio-based styrene has for the first time been successfully synthesized using a new metabolic pathway to produce this chemical that has previously only been produced from petroleum sources (McKenna and Nielsen, 2011). Styrene biosynthesis is currently limited by toxicity, but may have significant commercial implications. To be able to use the tool and quantify uncertainty in this pathway would display the potential of the tool for TEA and bio-based styrene.

With advancements in the development of new industrial biotechnologies, development of new techniques for evaluating the feasibility of these advancements is vital. These new techniques should mesh with the level of currently available data while building on previous industrial know-how. The ability to integrate these two items will allow better evaluation of new opportunities within this burgeoning technological space.

Objectives

The research objectives for this thesis were:

- ✧ Develop a spreadsheet-based economic model for sorbic acid via the CBiRC process
- ✧ Develop a general economic model for potential fermentation bioprocesses and validate between literature and current modeling techniques
- ✧ Evaluate the potential of bio-based styrene as a future bulk chemical from biorenewable resources using previously developed tool

Thesis Organization

This thesis contains a general introduction and literature review, three research articles, and a general conclusion. The introduction contains a general overview of the field of industrial technology, the objectives for this thesis, and the author's role in each paper.

For consistency and simplicity, all papers in this thesis use a citation style and subheadings appropriate to the flagship journal of the ASABE professional society; at the time of submission, each will be adjusted accordingly to the formatting requirements of the target journal. The first paper, *A Coarse Technoeconomic Model of a Combined Fermentation-Catalysis Route to Sorbic Acid*, is available as a meeting paper from the 2012 ASABE International Meeting. This paper examined and evaluated the overall complexity of approaching a commercially-viable bio-based sorbic acid process. The second paper, *Development of a Biorenewables Process Evaluation Tool: BioPET*, works through the development of an early-stage cost estimation tool for bio-based chemicals. The new tool, BioPET, is then validated against another commercial tool and literature values. The target journal for this paper is *Biofuels, Bioproducts & Biorefining*. The third paper, *Techno-economic Evaluation of Bio-Based Styrene from Escherichia coli*, is targeted to *The Journal of Industrial Microbiology*, and focuses on an early-stage TEA of bio-based styrene. This paper examines the uncertainty in production of styrene using commercial-scale biokinetic parameters and the economic feasibility and potential pitfalls of this new technology.

Author's Role

The primary author under advisement of the co-authors composed all of the papers presented in this thesis. The spreadsheets and tools were also developed by the primary author

under guidance from the major professor. The major professor suggested the approach of building a coarse TEA tool, and provided detailed editing of each of the manuscripts.

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CHAPTER 2: A COARSE TECHNOECONOMIC MODEL OF A COMBINED FERMENTATION-CATALYSIS ROUTE TO SORBIC ACID

Introduction

As petroleum continues to rise in price, bulk chemical production from biorenewable feedstocks becomes increasingly attractive. While bulk chemicals use less than 5% of a barrel of oil, they generate nearly 50% of the economic activity resulting from refining that barrel. This creates potential for the derivation of chemicals from biomass on an economic basis with greater ease than fuel because of the greater profit margin per unit, and because the total demand for carbon for chemicals is much lower than that for fuel. While the investigation for chemicals from biomass is nothing new (Cysewski and Wilke, 1978; Maiorella et al., 1984), new tools for improving biocatalysts have been steadily developed. These new tools (Jones and Kompala, 1999; Nevoigt, 2008) that serve metabolic engineers can enable rapid advancements in our knowledge base and a few (de Wit et al., 2010) have sought to predict where current research exists along the learning curve to a mature technology. Yet, little information exists regarding the production cost of bulk chemicals via these novel routes. While several projects have looked into the overall market potential of these biocatalyst-produced chemicals (Patel, 2006; Werpy and Petersen, 2004), the scope of the studies has been limited to chemicals that can be produced via biocatalysts only; no chemical catalysis step has been needed to bring these to market. Additionally, some scoping has been done on the direct catalysis of fructose to chemicals (Kazi et al., 2011), but it does not take advantage of producing upgraded intermediates from biocatalysis.

Recent technoeconomic work has focused on the short-term development of these chemicals, and on fuels (Jun et al., 2007; Song and Lee, 2006; Van Wegen et al., 1998), but there has been limited focus on single-site (Delhomme et al., 2009) or “one-pot”(Marr and Liu, 2011) processes where chemical upgrading of these products occurs at the same plant as the production of the intermediate. While the “one-pot” process focuses more on commodity chemicals rather than on bulk chemicals, the principles remain the same. The “one-pot” approach of combining biocatalysis and chemical catalysis into a single plant design can realize significant benefits over multiple unit operations both economically and from a lifecycle point of view. While the lifecycle analysis (LCA) must still be evaluated independently for each process, the Economic-Input-Output LCA (EIO-LCA) developed at Carnegie-Mellon University suggests that, in general, cheaper processes imply lower impacts (especially when changes in cost stay within the same sector – as with purchasing a smaller fermentation tank). The process uncertainties in the nascent industrial biotechnology sector likely present greater uncertainty than the EIO-LCA itself, so another reason to develop a coarse techno-economic assessment (TEA) of a sequential biological-chemical process is to help serve as a marker for economic and renewability improvements in the developing industry.

As with most economic ventures in the chemical industry, a first-pass, or coarse assessment, is needed to assess viability and identify limiting steps in the process, thereby guiding the development process. Researchers within the NSF Engineering Research Center for Biorenewable Chemicals (CBiRC) have identified and bench-validated several potential end-products from a sequential process train after examining plausible paths and connections between biologically producible compounds and chemical catalysis. One such product is sorbic acid, on which this coarse economic model is focused. Current sorbic acid production involves

the use of petrochemical feedstocks to produce a combination of sorbic acid and sorbate salts that both serve as anti-microbial agents in the food industry (Bohnet, 2003). Results from the coarse model are benchmarked against commercial sorbic acid prices of approximately $\$4.50\text{kg}^{-1}$, the commercial prices resulting from a process that utilizes ketene generation to polymer formation (Dorko et al., 2000).

Materials and Methods

The model developed for the purpose of evaluating the sequential sorbic acid production train incorporates key parameters from laboratory studies, such as yield, titer, selectivity, and conversion. These inputs fixed the parameterization of the model around which assumptions and best design practices were implemented to characterize the entire process. The model evaluates sorbic acid production in an optimistic manner for the purposes of providing insights into key bottleneck in the proposed process. In so doing, the model may provide early-stage feedback to guide future research and design of this process. No final purification process exists for sorbic acid for this potential route, but many have been speculated. Specifically, the butyl sorbate that comes from the catalysis step must be hydrolyzed and purified, but the design and economic evaluation of this stage is extremely difficult as no lab data exists. While hydrolysis remains a trivial step, by simple addition of water to push the equilibrium between the ester and the acid towards the acid, the sorbic acid partitions to the organic solvent (Dharmadhikari, 1992). To move the sorbic acid to the aqueous solution, salts have been used (Hans Fernholz, 1973) to increase crystallization efficiency. These final purification steps create a distillate stream containing water and n-butanol (hereafter referred to just as butanol), hydrolysis streams containing water with salts, butanol, and product, and eventually a product stream along with a wastewater stream. For cost estimation, the butanol lost to the butyl sorbate is non-trivial because

sorbic acid binds at a 1:1 molar ratio, and recovery of this butanol must occur via a recovery and recycle loop to be economically viable. Butanol can be recovered from water through decantation assuming that no catalysis steps are inhibited by butanol saturated with water; however with the presence of salts, separation is no longer trivial. Salts possess the ability to foul the catalyst, which cannot be overlooked in the design process. Overall uncertainty exists regarding purification techniques, catalyst fouling, and solvent choice, but these issues were not considered in detail in this work.

Processes

Fermentation

The process initiates with the microbial conversion of D-glucose to 4-Hydroxy-6-methyl-2-pyrone but is commonly and hereafter referred to as triacetic acid lactone, or TAL. We assumed that a strain of *S. cerevisiae* would be used as a biocatalyst, and that it would be realistic for such an organism to achieve a productivity, titer, and yield of $0.02 \text{ g} \cdot \text{L}^{-1} \cdot \text{hr}^{-1}$, $1.0 \text{ g} \cdot \text{L}^{-1}$, $0.10 \text{ g}_{\text{TAL}} \cdot \text{g}_{\text{glucose}}^{-1}$. Due to the organism's growth requirements, the media utilized is YEP complex media, although improvements are expected to reduce these requirements (DaSilva personal communication).

Separation

While separation has not been studied extensively, bench-scale separations have been conducted so that biologically-produced TAL can be subjected to subsequent catalysis (Dumesic personal communication). The bench methods used two adsorption columns to recover TAL and remove impurities from spent fermentation broth. While uncertainties exist regarding the full-scale implementation of extraction and separation methods, bench-scale proof of concept has been achieved.

Catalysis-Hydrogenation & Solid Acid

As with separations, details of the methods and current performance of the multi-stage catalyst system will not be presented here because of pending peer-reviewed publications. Current studies are investigating multiple catalysts to improve rate, selectivity, conversion, and non-fouling surfaces that present opportunities for significant economic and process improvements (Dumesic, personal communication).

Purification

As of this writing, no product has yet been purified, nor has the product sorbic acid been created. The butyl sorbate must be hydrolyzed, but upon hydrolysis, the sorbic acid remains primarily in the organic phase. A likely solution to this will incorporate distillation prior to hydrolysis and then exploit the temperature dependent solubility of sorbic acid in aqueous solution. Distillation of butyl sorbate in butanol will produce butanol in the distillate and butyl sorbate in the bottoms. Further investigation of this in lab scale studies will provide greater insight to the purification process.

Model

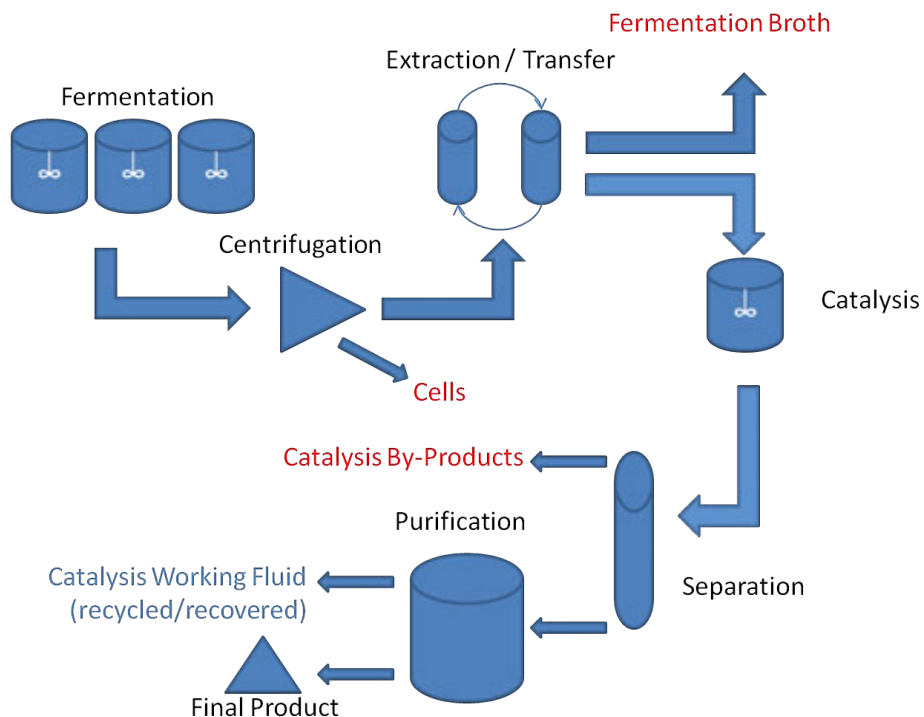


Figure 1: Simplified process diagram for proposed sorbic acid process and associated mass flows. Red lettering indicates mass flows leaving the system.

This model provided an estimate of, and insight into, the production cost of sorbic acid (reported on a \$/kg basis) from TAL via the process illustrated in Figure 1. To construct the process flow sheet Figure 1, the process assumed that fermentation is the first step in which TAL is synthesized. The broth from fermentation then undergoes a solids separation prior to loading into the adsorption columns. In the first column, the product is adsorbed and the remaining broth and constituents are sent to wastewater. The product is then desorbed into butanol, after which the butanol/TAL mixture flows through the second adsorption column for amino acid removal prior to catalysis. In catalysis, the solution flows through a hydrogenation reactor where TAL reacts to form HMTF. The butanol/HMTF mixture then enters two stages of solid acid catalysis where HMTF undergoes a condensation reaction and then a ring opening at a higher temperature,

resulting in butyl sorbate. Prior to the final transformation into sorbic acid, the butanol solvent must be removed. To remove the butanol, distillation is employed to exploit the large relative volatility between the solvent and solute molecules, with the butyl sorbate being the bottoms stream. The butyl sorbate is then mixed with water to achieve hydrolysis to sorbic acid. Due to its temperature-dependent solubility in water, sorbic acid can be purified from the aqueous solution by a crystallization step involving chilling the solution to ambient temperature with no use of refrigerants. The resulting sorbic acid crystals will then be dried and packaged. A great deal of uncertainty is associated with the purification steps following catalysis, as no lab studies have been conducted on this process, and this portion of the process is not modeled.

To approach the design, it was first determined that typical sizes of sorbic acid plants are presently in the range of 3,000 to 18,000 Mg per year. An annual production of 19,800 Mg per year (60,000 kg/day) was selected to constrain process flows, with this relatively large size chosen for the benefit of scale and to keep up with an anticipated market expansion due to sorbic acid outcompeting benzoic acid as a preservative used in food processing market. This annual flow rate was used to constrain the necessary design process to size components within the plant. To cost key unit operations, standard scaling laws were used, which correlate equipment cost to size (Peters et al., 2003) in Equation 1:

$$C_n = \frac{S_n}{S_o} \times C_o^n \quad (1)$$

Where:

C_n = new cost for newly sized piece of equipment

S_n = new size of equipment

S_o = size of equipment where previous cost data exists

C_o = cost of equipment where previous data exists

n = empirically-derived cost exponent

The two unit operations that utilize the exponent are the fermenters and the catalytic reactors which have exponents of 0.54 and 0.44 respectively (Peters et al., 2003).

Other unit operations less critical to the economics have been approximated in step changes and have been chosen at indicative sizes relevant to the base-case scenario (i.e. largest size available). The model computed upstream flows working backward from the assumed annual productivity, via assumptions about yield at each major unit operation. Yields of minor unit operations, such as centrifuges and pumps, are assumed to be 100%. The overall approach in this model computed major equipment costs (i.e., price paid to the manufacturer of the piece of equipment) for each unit operation, then to convert the total cost of all these pieces of equipment into a total plant cost via a Lang factor (Peters et al., 2003) which accounts for factors such as labor costs for installation, engineering expenses and construction overhead, and auxiliary facility costs. This establishes a total cost of construction for a new plant, complete with all major and minor facilities, on previously undeveloped land. Alternative methods can account for individual installation factors or delineate how much of the capital is directed towards individual plant construction processes; The Lang factor is a simplified aggregate of all these individual processes.

Overall Operating and Economic Assumptions

With capital costs computed as described above (i.e., determine total purchase cost of unit operations, convert to overall installed cost via Lang factor), total annual capital payments were computed assuming a 10 year, 10% internal rate of return. Overall plant operating time is needed to compute process flows, and annual operating hours and unscheduled downtime were assumed as shown in Table 2 below. Labor costs were estimated as a fraction of total capital invested (Table 1). While plant downtime and labor might be better estimated on the basis of single pieces of equipment (Peters et al., 2003), the uncertainty of the models inclusion of every piece of equipment, the actual operating hours for each piece of equipment, and amount of supervision required for each piece of equipment, was beyond the scope of this project.

Table 1: Key Economic and Plant Performance Parameters

Assumption	Value	Unit	Source
Internal Rate of Return (IRR)	10%	yr ⁻¹	(Kazi et al., 2010)
Operating Hours	7920	hrs	(Choi and Lee, 1997)
Unscheduled Downtime	1.1	hr _{total} /hr _{uptime}	(Van Wegen et al., 1998)
Operating Labor	10%	\$ yr ⁻¹ /\$ _{Capital}	(Peters et al., 2003)
Lang Factor	3.0	\$/ \$ _{Capital}	(Qureshi and Blaschek, 2000)

The operating-time parameters, along with annual production, guided the sizing of all the individual unit operations, thus providing baseline equipment costs, which were then converted into overall capital requirements as described previously. Overall capital requirements were then converted into annualized capital requirements via an amortization at the assumed IRR and 10-yr payout, while the labor factor in Table 1 was used to compute an annual labor cost.

Fermentation

Fermentation was calculated using Equation 2.

$$t_{fm} = \frac{t}{v} \quad (2)$$

Where:

t = fermentation titer (g/L)

v = production rate of product in fermentation (g/L/hr)

Fermentation time was amended by a downtime percentage, to account for time required for removal of broth, sterilization, and filling of new media. Fermenter working volume was assumed to be 80% of the total vessel volume.

The glucose loading per fermenter for the system was derived in Equation 3.

$$[glucose] = \frac{t}{Y_{ps}} \quad (3)$$

Where:

Y_{ps} = yield of product on substrate (kg_{product}/kg_{substrate})

Although CO₂ and cells production can be calculated, by-product economic value is not considered in this model. With fermentation producing the key intermediate necessary for all subsequent steps, the total number of annual batches was calculated using Equation 4.

$$N_b = \frac{N_d}{t_{fm}} \quad (4)$$

Where:

N_b = number of annual batches produced

N_d = number of days of plant operation (days)

t_{fm} = total time to complete a fermentation batch (days)

Knowing the number of batches run annually, the final titer (assumed previously), and the overall plant productivity, and correcting for any mass losses between fermentation and final product, the total working reactor volume was computed. Combining this with the maximum allowable fermenter working volume allowed computation of the required number of fermentation vessels. The cost of these vessels was found as described previously. Additional pieces of equipment such as stirring motors and compressors are sized according to Table 2 and the total capital cost associated with the entire fermentation step becomes the summation of all these expenses.

Table 2: Fermentation Parameters

Assumption	Value	Unit	Source
Corn Steep Liquor Inoculation	1%	w/w	(Rivas et al., 2004)
Maximum Fermenter Volume	3785	m ³	(Humbird et al., 2011)
Usable Fermenter Volume	80%	v/v	(Cysewski and Wilke, 1978)
Downtime Between Batches	20%	hr/hr	(Castilho et al., 2000)
Cell Mass Yield	$0.5 \times (1 - Y_{ps})$	g _{cells} /g _{substrate}	(Patel, 2006)
CO ₂ Mass Yield	$0.5 \times (1 - Y_{ps})$	g _{CO2} /g _{substrate}	(Patel, 2006)

Separation

As extensive lab data is not currently available on the adsorption process utilized for separation and the adsorption columns are similar to that of activated carbon, an activated carbon adsorption process was mimicked for a similar compound. The compound modeled was resorcinol as resorcinol is fairly closely related to TAL. Then from this extrapolation, modeling of the adsorption process can derive parameters from an activated carbon process using the Freundlich isotherm equation.

$$AR = K_f[A]^{1/n} \quad (5)$$

Where:

AR = adsorption ratio (kg product adsorbed/ Mg adsorbent)

K_f = Freundlich coefficient

$[A]$ = concentration of product in solution (kg/m^3)

n = Freundlich exponent

The absorption efficiency is then characterized by the exponential decay function that is then translated to a time needed to capture a certain amount of our product by solving the equation for adsorption duration:

$$AD = \tau \times (1 - \theta) \quad (6)$$

Where:

AD = adsorption duration

τ = adsorption time constant (hrs)

θ = the percent of product capture desired (%)

Table 3: Separation Parameters

Assumption	Value	Unit	Source
Freundlich Constant (K_f)	34.83	$\text{kg}_{\text{TAL}}/\text{Mg}_{\text{carbon}}$	(Kumar et al., 2003)
Freundlich Exponent ($1/n$)	0.23	dimensionless	(Kumar et al., 2003)
Adsorption Time Constant	16	hr^{-1}	(Kumar et al., 2003)
Apparent Density	0.977	Mg/m^3	(Kumar et al., 2003)
Max Column Size	628	m^3	(Peters et al., 2003)
Acetone Use	1%	$\text{kg}_{\text{acetone}}/\text{kg}_{\text{butanol}}$	guess

Significant uncertainty exists in this portion of the model, particularly in regard to the methionine content of the fermentation broth in a full-scale system, and these must be addressed in future feasibility studies as lab data becomes available. While both columns were modeled using data for resorcinol to mimic TAL, and the second column was exclusively for removal of methionine, this was our approximation of the process. Significant uncertainty exists in this portion of the model, particularly in regard to the methionine content of the fermentation broth in a full-scale system, and these must be addressed in future feasibility studies as lab data becomes available.

Catalysis-Hydrogenation

A concurrent plug flow reactor was assumed for the hydrogenation, similar to that used in glucose to sorbitol systems (James C. Chao, 1982). We assumed the capital and operating costs of this recovery loop to be negligible. Reactor size was computed by a combination of packing density, catalytic rate, and percent catalyst. Using the necessary volume of reactor to achieve a complete reaction, the volume was then translated to a large tube heat exchanger. Using an internal diameter of two inches and the max reactor size listed below, size and quantity of reactors could be determined per Equation 7 below:

$$V_{reactor} = \frac{\frac{kg_{TAL}}{day}}{\frac{kmol_{TAL}}{kg_{catalyst} \times min} \times \frac{1440min}{day} \times \frac{126.11g_{TAL}}{mol_{TAL}} \times \frac{kg_{catalyst}}{m^3_{reactor}}} \quad (7)$$

Table 4: Catalysis-Hydrogenation Parameters

Assumption	Value	Unit	Source
Packing Density	400	kg/m ³	(Merck, 2012)
Max Reactor Size	1000	m ²	(Peters et al., 2003)

Catalysis-Solid Acid

For solid acid catalysis, plug flow reactors were assumed, operating at moderate temperatures (80 – 200°C). With the large flow of daily solvent and low solubility of TAL in the initial butanol, both of these steps were sized based on hydraulic limitations provided by the manufacturer. The catalytic reactors were modeled as heat exchangers consisting of two inch internal diameter tubes, and costs were computed accordingly.

Table 5: Catalysis - Solid Acid Parameters

Assumption	Value	Unit	Source
Liquid Hourly Space Volume (LHSV)	8	m ³ _{liquid} /m ³ _{catalyst}	(Haas, 2005)
Max Reactor Size	1000	m ²	(Peters et al., 2003)

Purification

As little literature on butyl sorbate exists, the relative volatility for these two solutions was estimated by the boiling point method (Halvorsen and Skogestad, 2000). The resulting relative volatility is 13.4. The high relative volatility meant that a total of eight stages appeared to be optimal based upon basic distillation theory and MATLAB[®] code developed previously (Bequette, 1998). The resulting distillate was 99.91 mol/mol butanol, and the bottoms were 5.19% mol/mol butanol. To reach this bottoms purity, a reboil equivalent to the initial molar flow

rate of butanol was used. The result of the optimal case presented here used a reboil of 5.3 kmol/min.

Table 6: Purification Parameters

Assumption	Value	Unit	Source
Relative Volatility	13.4		(Halvorsen and Skogestad, 2000)
Equilibrium Stages	8	HETP	Calculated

Operating Costs

Operating costs are volatile, being dependent on factors such as corn yields, oil prices, and other market demands for grain and biomass. Feedstock prices were gathered from ICIS and literature (Peters et al., 2003), and are listed in Table 7.

Table 7: Prices of key process inputs

Assumption	Value	Unit	Source
Glucose	0.34	\$/kg	
Corn Steep Liquor	0.04	\$/kg	(Humbird et al., 2011)
Water	5.3	\$/1000kg	(Peters et al., 2003)
Acetone	1.39	\$/kg	ICIS
Hydrogen	2.00	\$/kg	(Doty, 2004)
Butanol	1.98	\$/kg	ICIS
Process Steam (790 kPa)	4.4	\$/1000kg	(Peters et al., 2003)
Electricity	0.06	\$/kWh	(Peters et al., 2003)
Wastewater	5.3	\$/1000kg	(Peters et al., 2003)

A sensitivity analysis was conducted for each case with a $\pm 5\%$ adjustment of every parameter (except for annual production) in the model. The results are reported in terms of percent change in minimum selling price (MSP), i.e., the sensitivity coefficients are reported for each of the parameters studied.

As the model aims to provide insight and early feedback, three scenarios were investigated to explore the impact of likely parameter changes and how the process economics

improved. The baseline scenario used estimates of readily attainable parameter values, while short-term and long-term scenarios examined process economics assuming medium and large (respectively) improvements in process parameters such as yield and catalyst life. The baseline scenario estimates how far current technology is from target economic viability, while the long-term scenario illustrates how the process economics might look once industrially viable performance parameters have been achieved, and can provide feedback on current design decisions. The short-term scenario illustrates how small changes in key parameters can provide large gains in economic viability. This intermediate scenario is important because it has the potential to speak to pilot-scale plants and startup companies looking to make the last jump from the short-term to long-term values. The scenarios are listed below with changes from one scenario to the next being highlighted in yellow.

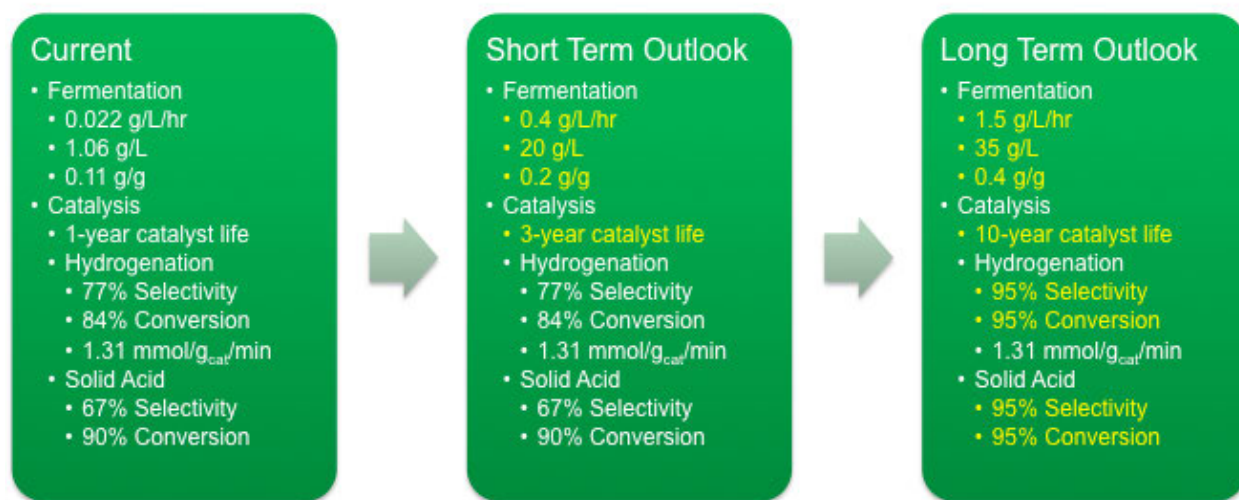


Figure 2: Summary of scenarios investigated in this work showing key parameter values.

Changes between scenarios are highlighted in yellow text.

Results

Results from each scenario were evaluated under three distinct sets of assumptions. In the non-ideal separation (NIS), non-ideal catalysis (NIC) case, the separation and catalysis yields were set to whatever value was used in the scenario. In the NIS & ideal catalysis (IC) case, catalysis yields were set to 100%. In the ideal separation (IS) and IC case, both separation and catalysis yields were set to 100%.

Current Case

In the current scenario, the largest portions of the cost come from fermentation and glucose. The difference between the NIS & NIC case and the IS & IC case illustrates the large cost of the yield losses in separation and catalysis. Figure 3 displays this descending trend from right to left in the various scenarios of ideal and non-ideal yields.

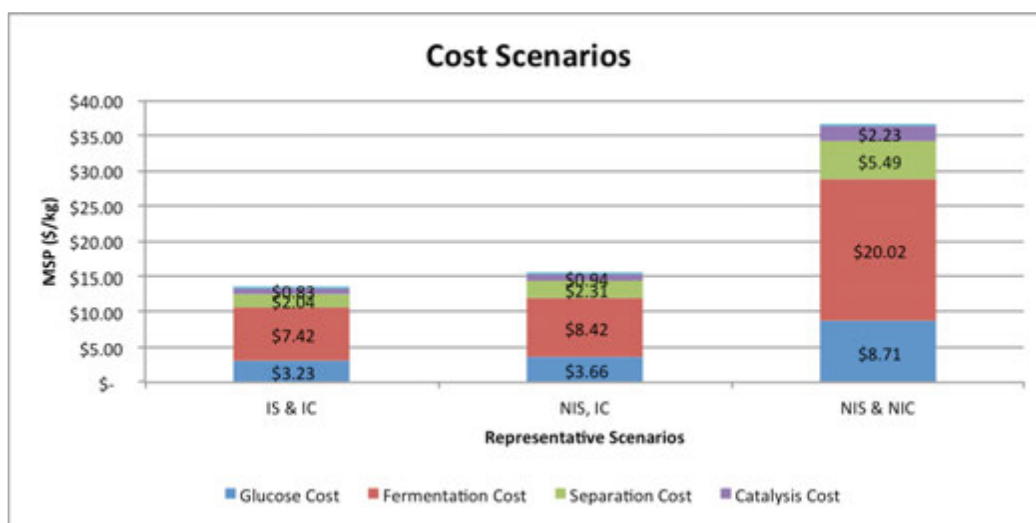


Figure 3: Cost Distributions for Three Scenarios in the Current Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.

Figure 4 provides a breakdown of the base cost into feedstock (glucose, fermentation, separation, and catalysis costs (each bar). Furthermore, each component cost is broken into the charges due to losses in fermentation (lowest segment of each bar), separation (middle segment of each bar), and catalysis (top segment of each bar). Figure 5 illustrates a trickledown effect, namely that as product is lost at each stage, not only does more product have to be made, but the entire scale of the project has to increase to account for such losses. Note that catalytic losses make up over 50% of all costs in the current scenario.

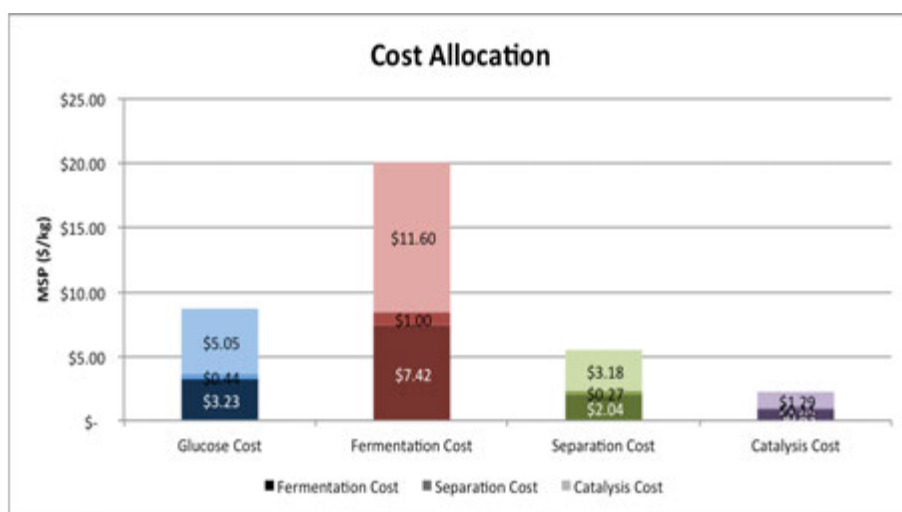


Figure 4: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the current case of sorbic acid

The top 10 most sensitive parameters are listed in Figure 5. A majority of these parameters are associated with fermentation due to the high cost of fermentation in the current case. The second sensitivity analysis, Figure 6, sets out to examine how current inputs are affecting the model. The largest of these appearing to be parameters affecting yields in catalysis, and apparent similar values for terms in fermentation with a slight tendency toward titer.

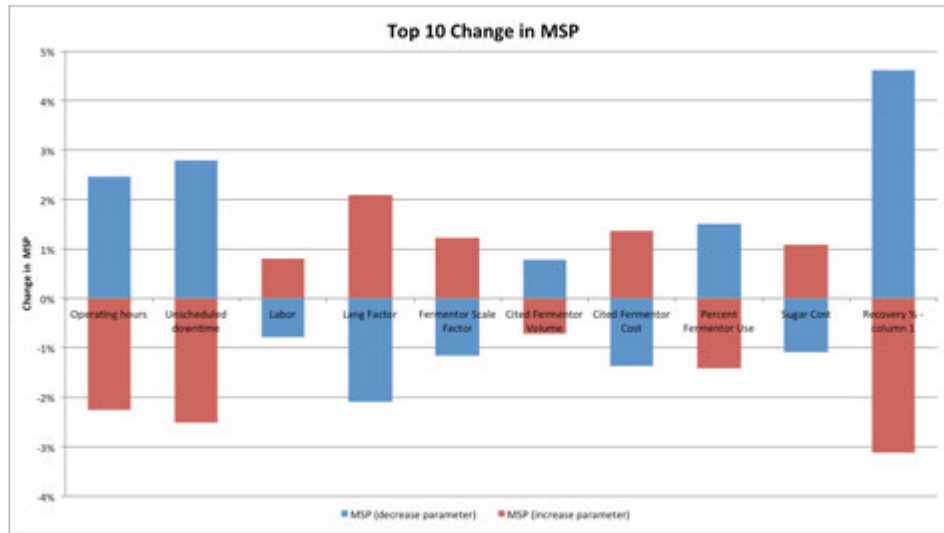


Figure 5: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the current case of sorbic acid

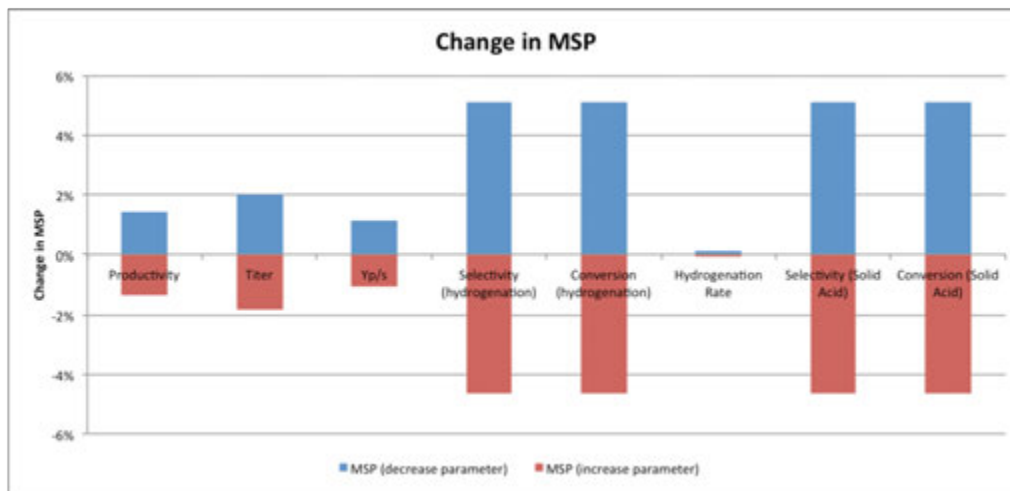


Figure 6: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the current case of sorbic acid

Short-term Case

This case takes an aim at improved inputs for fermentation due to its previously high cost in the current case. Resultantly, fermentation costs are significantly diminished with the greatest costs arising out of the feedstock and separation. However due to only improvements in the life

of the catalyst improving in the catalysis steps, over 50% of the NIS and NIC scenario still arises out of the yield losses due to catalysis. Figure 8 illustrates how the magnitude of this effect can inhibit the economic feasibility of the process and that if this effect can be ignored; the MSP drops below $\$5\text{kg}^{-1}$.

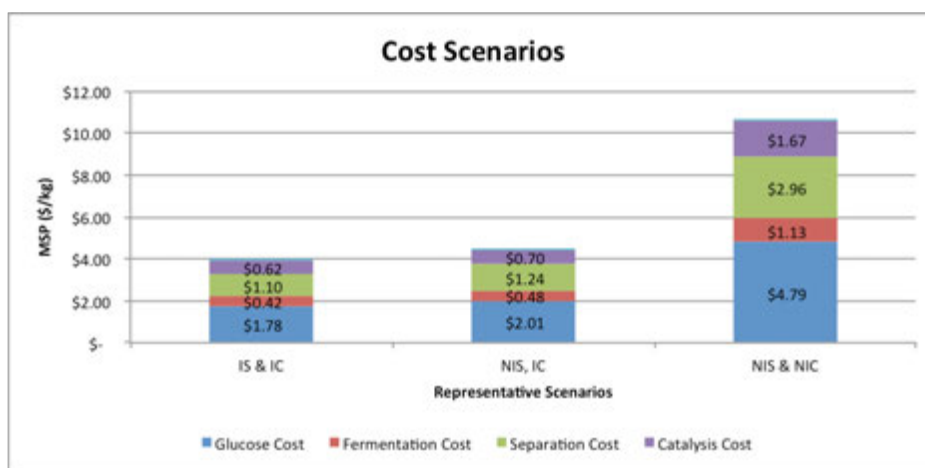


Figure 7: Cost Distributions for Three Scenarios in the Short-Term Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.

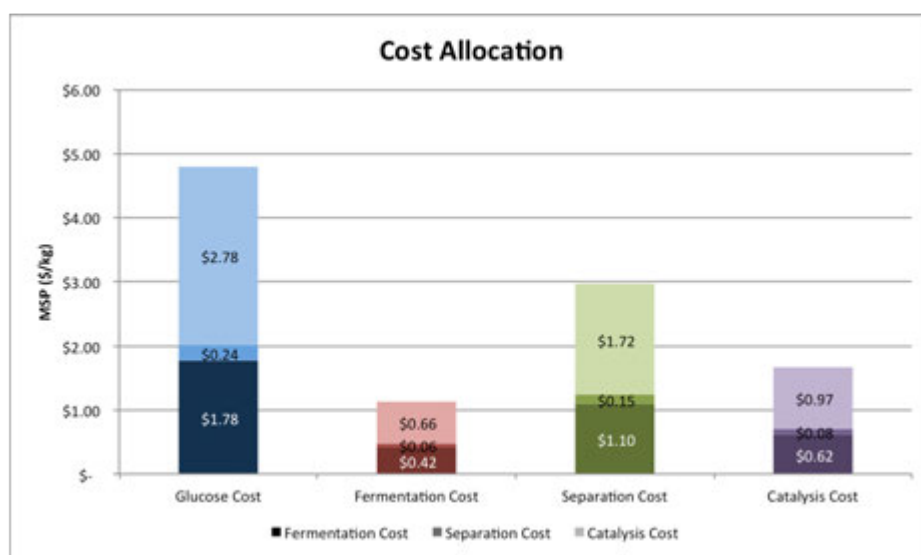


Figure 8: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the short-term case of sorbic acid

The sensitivity analysis for this scenario is notably different than the one conducted for the current scenario. The greater costs came out of separation and glucose, of to which now the model attributes a greater sensitivity. The TAL molecule's solubility in the solvent is critical in this scenario, and we also do not see as many of the key fermentation parameters as previously seen in the current scenario. Figure 10 continues this scheme as the yield losses in catalysis remain the most significant of the parameters followed by the yield of the product on glucose. Fermentation inputs have had a drastic shift, now more notably sensitive to yields on glucose than the productivity or titer

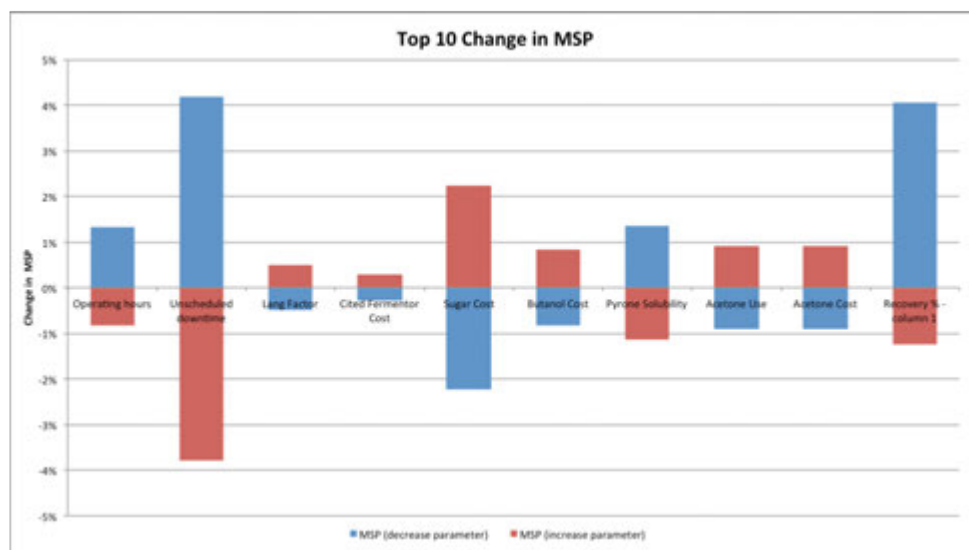


Figure 9: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the short-term case of sorbic acid

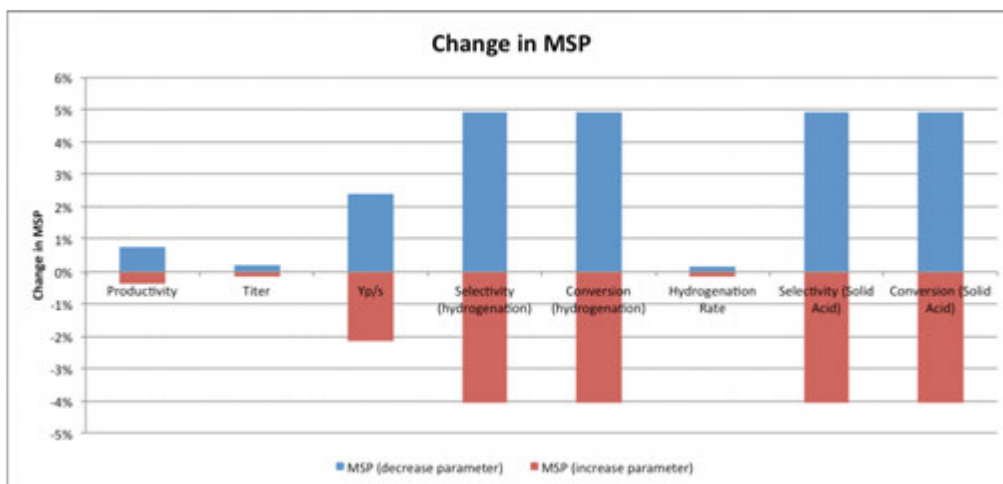


Figure 10: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the short-term case of sorbic acid

Long-term Case

With long term projections and following the sensitivity of our previous parameters, productivity was increased, titer was minimally increased, and yields were significantly increased in fermentation and catalysis. The lifespan of the catalyst was also increased for an all-around improvement in catalysis. These increases are demonstrated by Figure 11 and the minimal change between NIS and NIC to IS and IC; the largest costs occurring in the catalysis of TAL. No longer do the yield losses due to catalysis make up greater than 50% of the base cost, but rather, the cost results from the baseline fermentation.

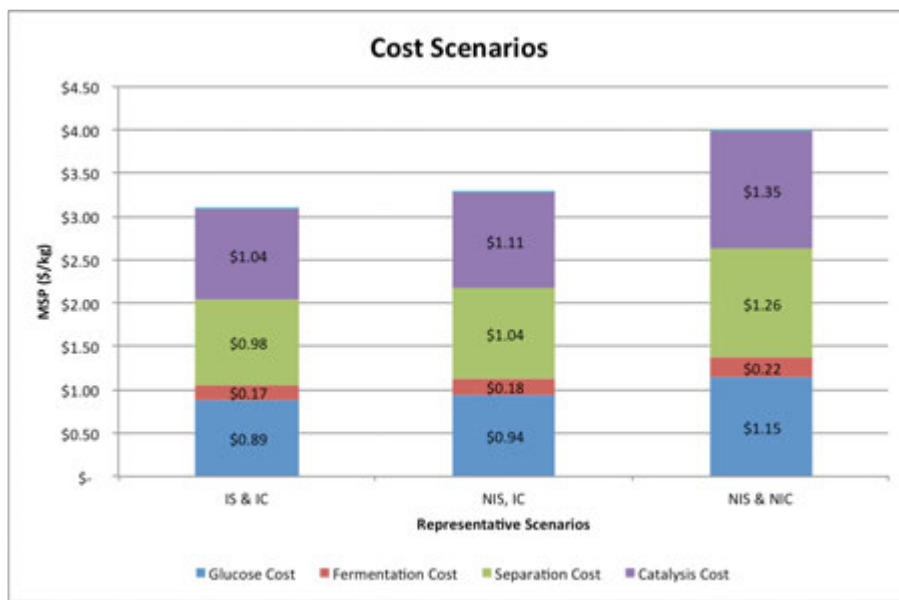


Figure 11: Cost Distributions for Three Scenarios in the Long-Term Case. IS = ideal separation, IC = ideal catalysis, NIS = non-ideal separation, NIC = non-ideal catalysis. See text for additional details.

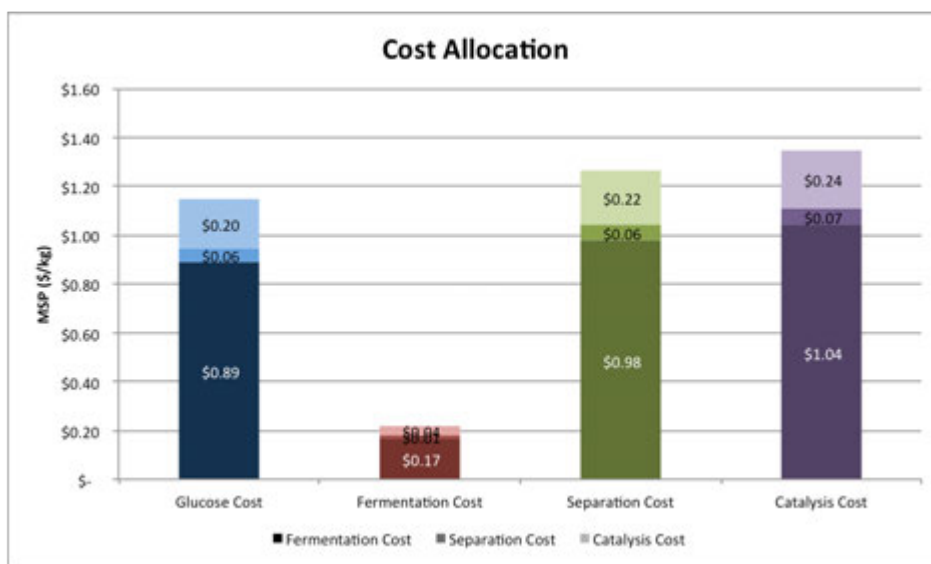


Figure 12: Respective distribution of costs due to yield losses out of fermentation, separation, and catalysis for the long-term case of sorbic acid

In the sensitivity analysis conducted for this case, costs are still arising mainly out of separation and therefore parameters regarding separation represent the greatest deal of

sensitivity. As productivity and titer have increased, their sensitivity is less than most of the top 10 parameters for this case. The yields for fermentation and catalysis will continue to remain sensitive through all cases because of the trickledown effect through all stages. This also should help describe the rest of the parameters and their sensitivity as the rest of the parameters must be less sensitive than time constant for adsorption at $\pm 0.14\%$ change in MSP.

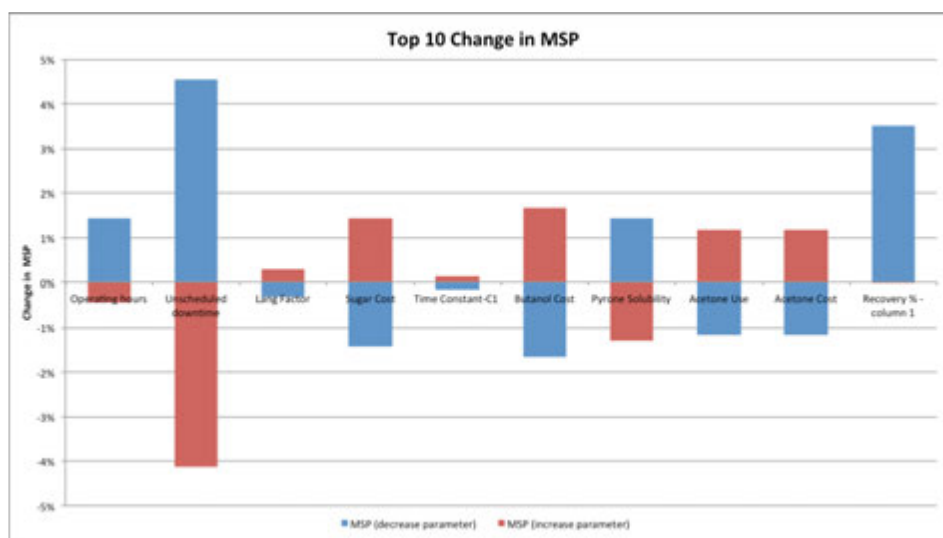


Figure 13: Sensitivity coefficients for model parameters given a $\pm 5\%$ change for the long-term case of sorbic acid

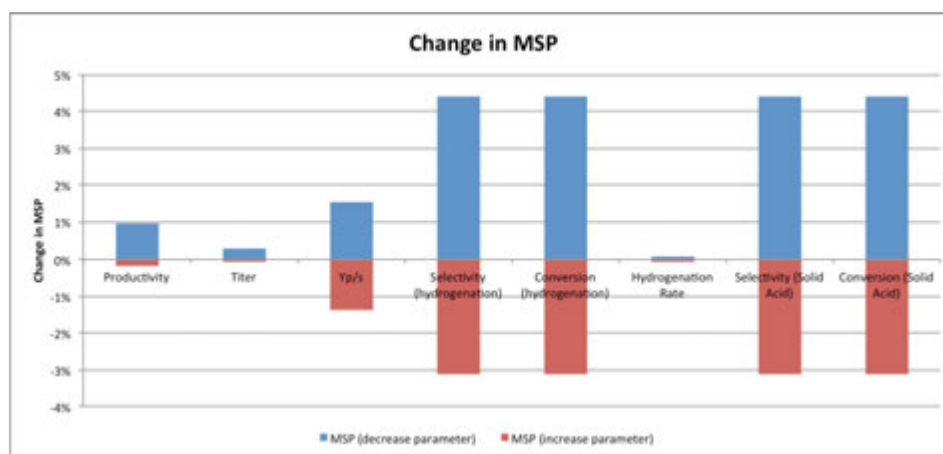


Figure 14: Sensitivity coefficients for model inputs given a $\pm 5\%$ change for the long-term case of sorbic acid

Discussion

With sorbic acid prices soaring at a maximum around \$4.50 kg⁻¹, this process becomes economically feasible in the long-term scenario with a best production price of \$3.27 kg⁻¹. This production price corresponds to the NIS, IC of the long-term scenario because the yield losses in catalysis can feasibly be recovered with continued research. Separation losses cannot actually be recovered because the design of adsorption columns using an isotherm requires knowledge about the desired amount of recovered product. Larger scale designs might incorporate breakthrough curves, but the limitation of knowledge around this subject prohibited that. With current lab-scale results, the process is far from commercialization with yield losses more than doubling the scale of the entire process, fermentation producing minimal amounts of TAL, and extremely slow production of TAL in fermentation. Improvements to these fermentation parameters can drastically reduce the base price as seen in the short-term scenario providing a quick jump to numbers not as far off from commercial prices. The progression to the long-term case increases yields such that the base-case scenario arrives within the realm of viability. However, additional capital costs will be incurred to provide a final purification of this product which may add more costs, but the recovery of butanol from the molecule of butyl sorbate will also recognize significant savings.

The current scenario suffers from large capital costs due to low productivity from fermentation. As productivity metrics improve, the cost decreases in a non-linear fashion. As separation appears to stand out in the short and long-term scenarios, further examination of these steps might provide the biggest insight into future technoeconomics of the same and similar processes. While the model only takes into consideration replacement of the activated carbon adsorbent once per year, where it has been suggested that there might actually be a correlation

between number of times the carbon is regenerated and carbon loss (Narbaitz and Benedek, 1983). Cost in catalysis can stand to be further examined but the costs are attributed to the production of butyl sorbate and less than $\$0.15 \text{ kg}^{-1}$ are a result of reactors and catalysts. It has been examined though that the lifespan of the catalysts are vital to this. With a life of 1-year, the catalyst cost increases to near $\$0.50 \text{ kg}^{-1}$. Ultimately another portion of significant cost is the use of acetone to wash and elute the amino acids off the columns. The amount of acetone use comes into question as the baseline cost in the long-term scenario is sensitive to its use and cost while very little data on the use of acetone exists. Further investigation of this process would benefit greatly from better understanding of activated carbon regeneration for the projection of long-term prices.

As the process improves, the inputs do not linearly transform the price, but rather follow an exponentially decay curve to a minimum achievable price. This indicates that while these small changes in inputs cause large shifts in the cost from the base-case, large improvements from the long-term case will see very minimal improvement in the base cost. This kind of relationship also validates not examining further cases beyond the identified long-term as the exponential relationship will only make minimal improvements in the process. However, it should be noted that these continued improvements will continue to improve the economics of such an endeavor, but tradeoffs between the investment into the research and the resulting improvement of the process should then be evaluated.

Conclusions

While final entry into the market will require overcoming some technological hurdles, this economic analysis suggests the potential feasibility of sorbic acid via TAL. Many uncertainties exist around this project, but the first pass analysis of this project presents the

plausibility of achieving MSP's of less than \$4 kg⁻¹. One study on the economic feasibility of this process will not be enough to confirm the viability projected in this study. Further studies should not only target improving the process, but confirming values that present the most sensitivity in the long-term scenario to decrease the risk associated with building a new chemical plant. The recommendation would also iteratively conduct economic feasibility analysis as the project continues to make progress to take into account the economic volatility of many projects, and bring to light new insights developed between each of the iterations. Economic analysis is not a one-pass process, but rather a continuous process that must undergo continuous revision from research and this is the first pass analysis of a potential route to sorbic acid.

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CHAPTER 3: DEVELOPMENT OF A BIORENEWABLES PROCESS

EVALUATION TOOL: BioPET

Introduction

The creation of chemical plants for bio-based chemicals represents an opportunity to produce value-added products from sugars. These chemicals represent an attractive alternative to biofuels because of their higher market prices compared to biofuels. In 2004, Werpy and Petersen identified ten chemicals that presented the greatest short-term opportunity for bio-based chemical production in the U.S., spurring tremendous efforts to increase the economic viability of these bio-based chemicals (Hermann and Patel, 2007; Jun et al., 2007; Sánchez et al., 2005; Song and Lee, 2006; Werpy and Petersen, 2004). Aggressive research and development efforts have led to increasing productivities and yields for these chemicals, but little is known about the economics of producing these bio-based chemicals at commercial scale.

Of the bio-based chemicals, ethanol provides a great segue to understanding the economics because of its large-scale deployment as a 1st-generation biofuel. Ethanol has been well studied with papers on topics ranging from process improvements, to technoeconomic analyses (TEAs), to life-cycle assessments, and can therefore provide a level of fundamental knowledge that may inform studies about future bio-based chemical opportunities (Kazi et al., 2010; Kwiatkowski et al., 2006; Mangat et al., 2010; Michael et al., 2007). Robust TEA's, in particular, have the ability to illuminate process bottlenecks and to clarify how process alternatives will impact the production costs. These TEA's require extensive knowledge of process parameters and design details only available during the latter stages of a project after

many years of research have illuminated such facts. However, early-stage cost estimation is critical to helping companies and applied academic research centers chart a course through translational research and towards economic viability.

As engineers develop and enhance our ability to convert sugar into chemicals via metabolic pathways (Nielsen, 2001), (Bozell and Petersen, 2010) or with novel hybrid fermentative-catalytic processes in advanced biorefineries (Nikolau et al., 2008); comprehensive and accurate data of these operations at significant scale will be years away. And yet, strong evidence regarding the economic viability of a particular chemical is needed early in the process to warrant investment of time and money. Using simplified estimations, one can provide an early-stage TEA. Strong TEA capabilities exist commercially in tools such as Aspen Economic Analyzer™ and Intelligen SuperPro Designer®, both of which provide estimations of capital and operating costs. These tools also require a level of detail that is typically unavailable at early stages in process evaluation. While preliminary cost evaluation methods have been outlined by several authors (Peters et al., 2003; Turton et al., 2010), we are unaware of any widely-available early-stage TEA model or tool for bio-based chemicals. To address this gap, we developed a spreadsheet-based model to provide early-stage TEAs of bio-based chemicals, named BioPET (Biorenewables Process Evaluation Tool). Key criteria used in the development of BioPET were ease of use and minimal data inputs for process evaluations. To operate the tool, users need a basic knowledge (or educated guesses) for each unit operation comprising their overall process design of interest. Once developed, BioPET was compared against SuperPro Designer® and results from the BREW project for a suite of three chemicals: ethanol, succinic acid, and adipic acid (Patel, 2006).

Methods

In keeping with the vision of the NSF Engineering Research Center for Biorenewable Chemicals (CBiRC), the model focuses on combined fermentative-catalytic processes by assuming a directly fermentable source regardless of initial feedstock that can accommodate any process described within the CBiRC area of research pictured in Figure 1.

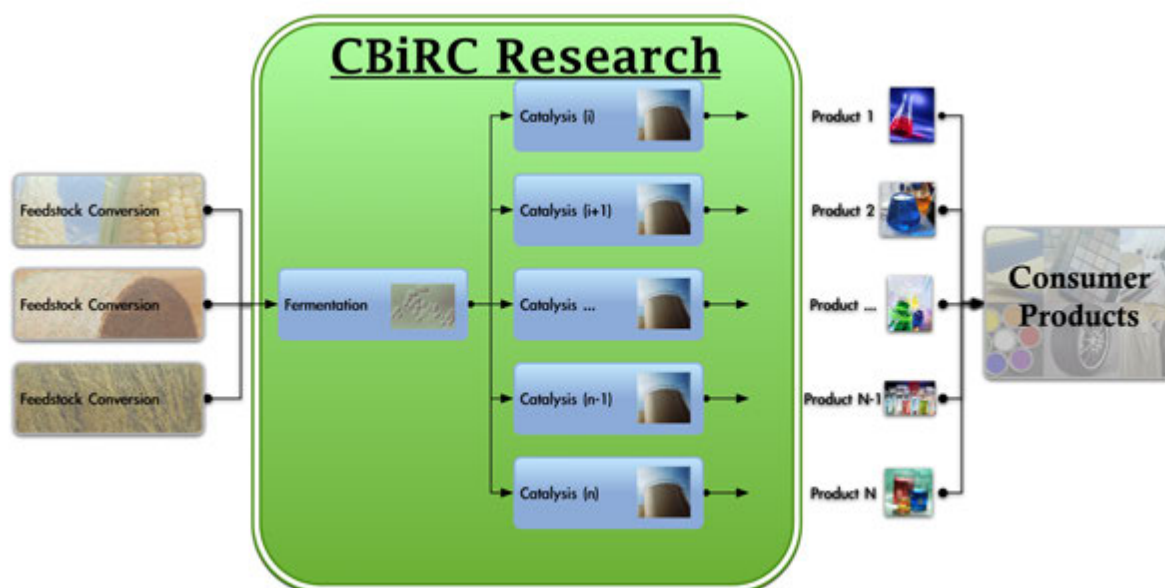


Figure 1: CBiRC's focus in chemical commodity value chain

It is noteworthy that the model does not consider upstream processes such as starch hydrolysis or pretreatment and hydrolysis of lignocellulosic biomass – these were considered outside the scope of the model and the feedstock costs were considered as a lumped parameter to include the costs of the initial source and conversion technology. BioPET assumes the following carbon flow: fermentation, followed by a separation stage, followed by up to three catalytic processes, finishing with up to two purification stages. All stages can be toggled on and off to allow for process flexibility. Due to the inherent complexity of the separation, catalysis, and purification processes, an approach was taken to accommodate this complexity while allowing

for a relatively simple user interface: First, the separation and purification choices in BioPET are limited to a menu of 2 – 4 items as shown in Table 1. Second, BioPET incorporates multiple assumptions about each separation and catalytic unit operation that allow for minimal input from the user. Finally, BioPET only considers a stream consisting of a primary product and solvent. This binary system uses mass balance equations and relationships to characterize all steps post-fermentation.

Table 1: Separation and Purification Unit Operations

Separation	Primary Purification	Secondary Purification
Adsorption	Adsorption	Crystallization
Distillation	Crystallization	Distillation
In-Situ	Distillation	
Liquid-Liquid Extraction	Liquid-Liquid Extraction	

A list of the key inputs is listed in Table 2. Using these inputs, and the assumptions and equations described in the following sections, process cost estimations can be made.

Table 2: Input Variables Necessary for Unit Operations

Unit Operation	Input Variable	Input Variable	Input Variable	Input Variable
Fermentation	Productivity	Titer	Yield	-
Adsorption	Freundlich Coefficient	Freundlich Exponent	Yield	-
Crystallization	Mass Separation (Y/N)	Purity	Yield	Separating Agent Use
Distillation	Relative Volatility	Purity	Yield	-
Liquid-Liquid	Distribution Coefficient	Fraction in Extract	Yield	-
Catalysis	Solid Acid/Base	Selectivity	Conversion	Commercial Resin
	Hydrogenation			Ni-Raney
				Platinum
				Palladium

Unless otherwise specified, a standard scale-factor approach, as embodied in Equation 1 (Peters et al., 2003), was used to adjust capital costs based on unit operation size.

$$C_n = \frac{S_n}{S_o} \times C_o^n \quad (1)$$

Where:

C_n = new cost for newly sized piece of equipment

S_n = new size of equipment

S_o = size of equipment where previous cost data exists

C_o = cost of equipment where previous data exists

n = empirically-derived cost exponent

Individual exponents are listed for each component (Peters et al., 2003) in the following sections, and all materials of construction are assumed to be stainless steel grade 304 to account for the reactivity of biological molecules.

Fermentation

Using productivity, titer, and yield, BioPET computes baseline fermentation time and sugar demand. The baseline fermentation time is then increased by 20% to account for downtime needed for vessel emptying, cleaning, and refilling. The required fermentation volume is computed based on the required mass flow of product and on the downtime-corrected productivity. The number of equal-volume fermenters necessary is then calculated; each primary fermenter is also associated with a seed fermenter having a volume that is 10% of the primary fermenter.

Table 3: Fermenter Assumptions

Maximum Size	3785 m ³	(Humbird et al., 2011)
Usable Percentage	80% m ³ _{used} / m ³ _{purchased}	(Cysewski and Wilke, 1978)
Capital cost of 757 m³ Vessel	590,000 \$	(Humbird et al., 2011)
Cost Exponent	0.54 Dimensionless	(Humbird et al., 2011)
Downtime	20% Downtime/Uptime	(Castilho et al., 2000)

Following the series of Equations 2-5, the number of fermenters and their respective size can be calculated.

$$V_{fm} = \frac{m_T}{(c_{fm})(FV_{fm})(\eta_T)(MR)} \quad (2)$$

Where:

V_{fm} = annual volume of fermentation media (m³)

m_T = annual production of fermentation product (kg)

c_{fm} = final titer of fermentation product (kg/m³)

FV_{fm} = useable fraction of fermenter volume (dimensionless, purchased volume/usable volume)

η_T = mass conversion efficiency (dimensionless, kg final prod./kg fermentation prod.)

MR = mass ratio (dimensionless, molecular weight of final product/molecular weight of molecule produced in fermentation)

$$N_b = \frac{N_d}{t_{fm}} \quad (3)$$

Where:

N_b = number of annual batches produced

N_d = number of days of plant operation (days)

t_{fm} = total time to complete a fermentation batch (days)

$$N_{fm} = \text{Round Up} \left(\frac{V_{fm}}{N_b \times V_{max}} \right) \quad (4)$$

Where:

N_{fm} = number of fermenters required

V_{max} = maximum attainable volume in a purchased fermenter (m^3)

$$S_{fm} = \frac{V_{fm}}{N_b \times N_f} \quad (5)$$

Where:

S_{fm} = equally-sized volume of the fermenters (m^3)

The media sugar requirement is back-calculated using the input titers and predicted yield.

Additional nutrients are supplied in the form of corn steep liquor, a common supplement for various micronutrients, and are added at a rate of 1% w/w.

Centrifugation

The model includes a centrifugation stage immediately downstream of fermentation to account for removal of cell mass from the broth, which cannot be deselected. All feedstock that does not end up in the final fermentation product is assumed to be converted into cell mass and metabolic by-products that are then combined into a single stream and separated out.

Centrifugation energy requirements and size are approximated for a disc-stack centrifuge using Table 4.

Table 4: Centrifuge Assumptions

Volumetric Energy Requirement	1.2 kWh/m ³ _{throughput}	(Peters et al., 2003)
Maximum Centrifuge Size	400 kW	(Peters et al., 2003)
75.5 kW Centrifuge	140,400 \$	(Peters et al., 2003)
Cost Exponent	0.67 Dimensionless	(Peters et al., 2003)

$$N_c = \frac{FV_{fm} \times S_{fm}}{t_d} \times \frac{VE_c}{V_m} \quad (6)$$

Where:

N_c = number of centrifuges

t_d = fermentation downtime (hrs, time used for cleaning and refilling)

VE_c = Centrifuge volumetric energy requirement (kWh/m³ fermentation liquid)

V_m = Maximum attainable size of a centrifuge

Adsorption

Adsorption is typically used to remove dilute contaminants or products from a stream – and large-scale adsorption is primarily focused on pollutant removal rather than product recovery. When increasingly large adsorption column volumes are needed, we assumed they are achieved by adding multiple additional units, not by simply making a single unit larger. The number of columns computed in BioPET is done so with non-integer results allowed (e.g., a plant could have 8.6 columns).

Table 5: Adsorption Assumptions

Time Constant	16 hrs ⁻¹	(Kumar et al., 2003)
Apparent Density of Adsorbent	0.977 Mg/m ³	(Kumar et al., 2003)
Maximum Size of Column	628 m ³	(Peters et al., 2003)
628 m³ Column	109,200 \$	(Peters et al., 2003)

While exotic resins can be used for adsorption systems, at large scale and high throughput, their upfront costs may not exceed their benefits. Activated carbon is frequently used in large-scale adsorption systems; the cost of the adsorbent in the model therefore follows that of activated carbon with a cost of \$1.00/kg. To calculate the quantity of adsorbent required, Equations 7-8 are used.

$$AR = K_f[A]^{1/n} \quad (7)$$

Where:

AR = adsorption ratio (kg product adsorbed/ Mg adsorbent)

K_f = Freundlich coefficient

[A] = concentration of product in solution (kg/m³)

n = Freundlich exponent

$$N_A = \frac{-[\tau_A \times \ln(1-Y_A)] \times m_p}{Y_A \times AR \times t_f} \quad (8)$$

Where:

N_A = adsorbent needed (Mg)

τ_A = adsorption time constant (hrs⁻¹)

Y_A = adsorption yield

m_p = mass of product per batch (kg)

Adsorption columns use various techniques to provide separation such as size, charge, and specialized ligands. A majority of these techniques requires adsorbent regeneration or cleaning similar to that of activated carbon and commercial resins (Haas, 2008). As this regenerating solution imparts activity back to the adsorbent, the solution is considered to be consumed annually at a rate of twice the sum of the volume of all adsorption columns.

Crystallization

Crystallization is a viable separation technique for several of the compounds evaluated by both the BREW project and the United States Department of Energy, and potentially many more bio-based chemicals (Patel, 2006; Werpy and Petersen, 2004). Several techniques in particular exist for producing crystals from their respective solution. Two of these techniques rely on steam for cooling or evaporation of the solvent to generate a saturated solution from whence the crystals can then be separated. Another method relies on the addition of a mass-separating agent to the solution to precipitate the crystal of interest out. This method utilizes less energy but often produces a secondary product at the expense of the mass-separating agent. Crystallizers come in such drastically different configurations to provide different levels of separation based on product requirements that often a highly specific crystallizer design must be implemented (Jones, 2002). As this highly specialized process does not easily lend itself to process design, an external forced circulation crystallizer was chosen as an optimistic choice because of its ability to run continuously and at a high production rate, typically between 5000 kg hr^{-1} and $50,000 \text{ kg hr}^{-1}$ (Walas, 1990). The cost of external forced circulation crystallizers follows Equation 9.

$$\$_c = 4e^{[4.868 + .3092 \ln X + .0548(\ln X)^2]} \text{ (Walas, 1990)} \quad (9)$$

Where:

$\$_c$ = cost of crystallizer

X = flow rate of crystals (klb hr^{-1})

Table 6: Crystallization Assumptions

Maximum Crystallizer Size	50,000 kg hr^{-1}	(Walas, 1990)
Dryer 408 $\text{kg}_{\text{product}}/\text{hr}$	158,000 \$	(Genskow, 2008)
Increase in Dryer Cost/$\text{kg}_{\text{product}}$	112 \$ kg^{-1}	Unpublished Analysis
$\text{kg}_{\text{steam}}/\text{kg}_{\text{product}}$	2/4 kg kg^{-1}	

While crystallization can produce very pure products, often the crystals must still be purified from the solvent. Dryers are required to finish separation and purification of the product. An examination of dryers from *Perry's Chemical Engineering Handbook* (Genskow, 2008) provided a cost and a linear relationship (data not shown) between crystal production and cost of 112 $\$/\text{kg}_{\text{production}}$. The steam requirements were set to 2 $\text{kg}_{\text{steam}}/\text{kg}_{\text{product}}$ for a process using precipitation and 4 $\text{kg}_{\text{steam}}/\text{kg}_{\text{product}}$ for temperature-sensitive crystallization. These values were multiplied by the inverse of the purity of the crystallization step to obtain annual steam requirements.

Distillation

Distillation represents a well-characterized process unit operation within chemical engineering and can be estimated using the Fenske-Underwood sizing calculations as represented by Equation 10 (Peters et al., 2003). Fenske-Underwood assumes a constant relative volatility to construct the necessary number of equilibrium stages. Using Table 7 for costs and combining it with Equations 10 - 11, a distillation tower cost can be calculated.

Table 7: Distillation Assumptions

Murphee's Tray Efficiency	50%	Dimensionless	
Distance Between Trays	0.5	m	
Maximum Column Height	55	m	
10m Column	195,000	\$	
Cost Exponent	0.62	Dimensionless	
Sieve Tray Cost	3900	\$	
Tray Multiplier	1.25	<30 Trays	
	0.98	30≤n≤60 Trays	
	0.97	>60 Trays	

$$N_{\min} = \frac{\log \left[\left(\frac{X_d}{X_b} \right) \left(\frac{Y_b}{Y_d} \right) \right]}{\log(\alpha)} \quad (10)$$

Where:

N_{\min} = minimum number of stages

X_b = fraction of product in the bottoms

X_d = fraction of product in distillate

Y_b = fraction of solute in bottoms

Y_d = fraction of solute in distillate

α = relative volatility of product and distillate

$$N_{actual} = \frac{N_{\min}}{\varepsilon_{tray}} \quad (11)$$

Where:

N_{actual} = actual number of sieve trays required

ε_{tray} = Murphee tray efficiency

While these equations can be used to predict the tower capital costs, these equations cannot account for the energy requirements of the condenser and reboilers. Nor can condensers and reboilers be sized without further process details. It can be theorized that the annual cost of steam ranges from 137% to 191% the amortized cost of the column (Kookos, 2003). A value of 150% was chosen for BioPET. The capital cost of the heat exchanger can be approximated similarly as approximately 100% of the capital cost of the column bare module (Kookos, 2003).

Liquid-Liquid Extraction

Liquid-liquid extraction represents a separation method applicable to potential fermentation products such as succinic acid (Kurzrock and Weuster-Botz, 2010). Utilizing the Kremser assumptions and the necessary inputs as listed in Table 8, a calculation using Equations 12 and 13 (Albright, 2009) can calculate the necessary sizing requirement of an extraction column.

Table 8: Liquid-Liquid Extraction Assumptions

Murphee's Tray Efficiency	10%	Dimensionless	
Distance Between Stages	0.5	m	
Maximum Column Height	55	m	
10m Column	195,000	\$	
Cost Exponent	0.62	Dimensionless	
Sieve Tray Cost	3900	\$	
Tray Multiplier	1.25	<30 Trays	
	0.98	30≤n≤60 Trays	
	0.97	>60 Trays	

$$E_e = \frac{K_{Dt}E}{R} \quad (12)$$

Where:

E_e = extraction factor

K_{Di} = distribution coefficient of the product in its respective solvent

E = extract flow rate

R = raffinate flow rate

$$N_{min} = \frac{\ln\left[\frac{(E+\phi_e-1)}{\phi_e}\right]}{\ln E_e} - 1 \quad (13)$$

Where:

ϕ_e = fraction of product not extracted

After the extraction column has been sized, it is then adjusted using an efficiency calculation as described by Equation 11.

Catalysis

The chemical reaction pathways for bio-based chemicals are typically more temperature-sensitive than those of petrochemical pathways (Chia et al., 2012). This temperature sensitive property implies widespread use of isothermal packed-bed reactors for catalysis. To model these reactors, a large tube heat exchanger is used. A standard 2 in. tube was used so that the available internal packing volume was 102 m²/m³ of available heat exchange area. A solvent density of 810 kg/m³, typical of many organic solvents, such as n-butanol, was applied if the separation step prior was adsorption; otherwise the solvent density was that of water. In the packing of catalytic reactors, several common catalysts were chosen for different types of reactions: hydrogenation, solid acid, and solid base. As solid acid and solid base catalysts are both resin-based, all associated properties were considered identical to one another. An analysis of Amberlyst™ catalysts of both solid acid and solid base was conducted and no significance in differences of

packing density or LHSV was discovered. A constant liquid-hourly-space-volume, LHSV, was assumed for all catalysts and packing density was varied by catalyst as follows: solid acid/base - 719.5 kg/m³, 5% platinum - 810 kg/m³, 10% palladium - 400 kg/m³ –, 90% nickel/raney -1200 kg/m³.

Table 9: Catalysis Assumptions

Solvent Density	810 kg/m ³	
Bed Volumes (LHSV)	5.2 m ³ _{solution} / m ³ _{resin}	Unpublished Analysis
Percent Usable Reactor	70% Dimensionless	
Maximum Surface Area	1000 m ²	(Peters et al., 2003)
880 m² Reactor	149,500 \$	(Peters et al., 2003)
Cost Exponent	0.44 Dimensionless	(Peters et al., 2003)

Cost

Chemical plants incur two significant types of costs; capital and operating costs. Often operating expenses can dominate the total cost of production, such as utilities and feedstocks, accounting for greater than 75% of total manufacturing costs (Cysewski and Wilke, 1978). To account for these costs and other major pieces of specific unit operations, such as the metal for a catalyst, a list of assumed costs is described in Table 10. These, while potentially overgeneralizing, provide a basis for evaluating tradeoffs of processes under identical assumptions.

Table 10: Cost of Key Materials and Supplies

	\$/kg
Plant Utilities	
Water	0.00053
Electricity*	0.06
Wastewater	0.00053
Steam	0.0044
Fermentation	
Feedstock	0.15
Corn Steep Liquor	0.2
Separation	
Activated Carbon	1
Commercial Resin	15
Solvent	1
Extractant	1
Resin Regenerating Solution	0.3
Catalysts	
Ni-Raney	40
Platinum	50000
Palladium	22600
Commercial Resin	10

*Unit is \$ kWh⁻¹

Model Comparison: Approach

SuperPro Designer[®] is a chemical process simulation program capable of providing detailed information on process design. SuperPro was used as a benchmark for BioPET for evaluating BioPET's performance. However, without knowing all the process details or knowing the source of feedstock, a simplified SuperPro model must be used. An existing ethanol plant model (Kwiatkowski et al., 2006) was used as comparison to the model displayed in Figure 2. These costs were distributed near identical with an increase in utilities cost of the simplified model due to the lack of heat integration and plant-wide pinch analysis. Pinch analysis has previously shown approximately 30% reduction in utilities, and accounting for this magnitude of savings appears to bring these values into proximity of one another (Kemp, 2007; Khan and Riverol, 2007). With simplified models being a modest representation of their detailed

counterparts, simplified models were then evaluated in SuperPro Designer[®] and compared to BioPET using the following chemicals: ethanol, succinic acid, and adipic acid.

While SuperPro provides a level of comparison that is necessary for validation of a simulation tool, this provides limited validation of final product cost. To provide another benchmark for cost estimation, the BREW project was used in accordance with its assumptions to provide a cost-comparison of each product (Patel, 2006). This allows for all comparisons to be conducted both over a range of chemicals and fermentable sugar cost. BREW will measure the accuracy while SuperPro analyzes the precision, and between both they will provide a foundation for evaluating BioPET as an early-stage cost-estimation tool for biorenewable processes.

Model Comparison: Ethanol

Ethanol represents a well-studied process with corn ethanol representing a bulk of the studies and is often used in the fuels industry but can also see application in chemical synthesis (Cysewski and Wilke, 1978; Hamelinck et al., 2005; Maiorella et al., 1984). Ethanol is converted in a one-step fermentation process and then run into a series of distillation columns to provide pure ethanol as seen in Figure 2. The ethanol process designed is to produce 40 million gallons per year (MGY) with 5% denaturant to be added prior to the final product (Kwiatkowski et al., 2006). The process takes an input of yeast, glucose, and water to the fermenter and is aerated at 0.20 volumes of air per volume of liquid (at STD). This is converted into 100 g/L ethanol and yeast. The liquid is sent to a holding tank and then on through to a heater and the first of the distillation columns. Using a series of two columns to remove a majority of water, a molecular sieve removes the remaining water to content of <0.5% g/g. The ethanol is finally sent onto a holding tank to be blended with denaturant and sold.

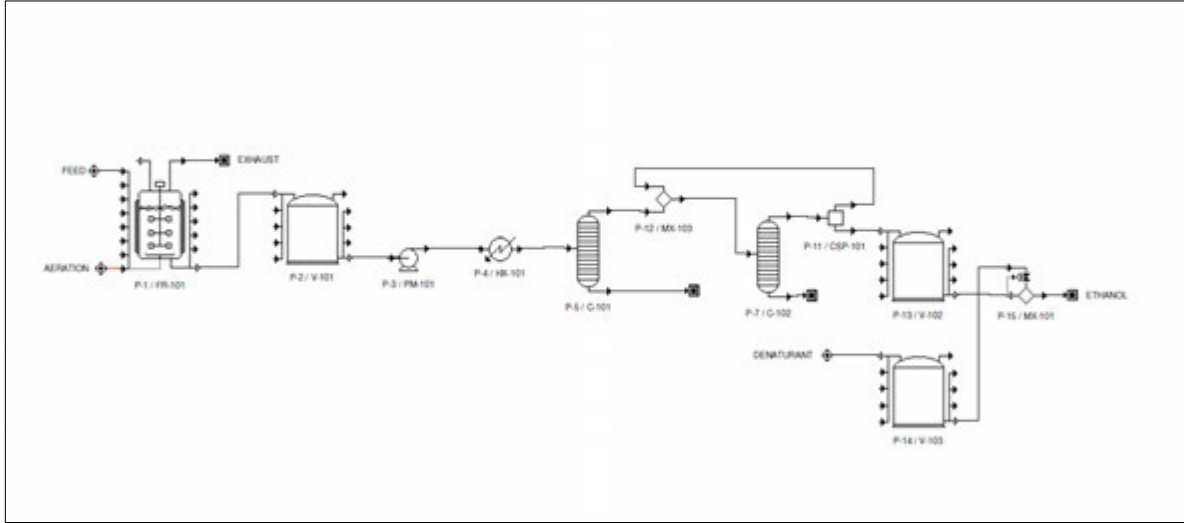


Figure 2: SuperPro process flow diagram for ethanol

BioPET uses a spreadsheet and effectively mimics basic process flow diagrams (PFDs). Basic inputs are required of the user as suggested in Table 2, and are described for the simulated ethanol scenario in Table 11.

Table 11: BioPET Inputs – Ethanol

Input Variable	Value	Source
Production Size	40 MGY (119.1 KTA)	(Kwiatkowski et al., 2006)
Operating Days	330 days	
Internal Rate of Return	10%	
Plant Operating Life	10 years	
Mass Ratio	1	
Fermentation		
Productivity	2 g/L/hr	
Titer	100 g/L	
Yield	0.51 g/g	Max. Theor. Yield
Separation	Distillation	
Relative Volatility	10	(Kwiatkowski et al., 2006)
Purity	0.5	(Kwiatkowski et al., 2006)
Yield	99.9%	(Kwiatkowski et al., 2006)
Catalysis A → C	N/A	
Primary Purification	Distillation	
Relative Volatility	10	
Purity	0.95	Azeotropic Mixture
Yield	99.9%	
Secondary Purification	N/A	

Model Comparison: Succinic Acid

Succinic acid is considered to be a potential new source for polymeric material (Werpy and Petersen, 2004) and is in the process of being commercialized by several companies such as Myriant and BioAmber each of who is constructing bio-based succinic acid plants at present. Succinic acid represents many opportunities for chemical intermediates that serve bulk chemicals, food additives, and pharmaceuticals (Delhomme et al., 2009; Song and Lee, 2006). By re-engineering *E. coli*, the highest yields of succinic acid were obtained and approach 1.6 mol/mol of succinic acid per mole glucose (Sánchez et al., 2006). While ethanol uses a Lang Factor of 3, more complex operations will consist of increased cost and a more conservative estimates of this might use a Lang Factor of 6 as was chosen for succinic acid (Peters et al., 2003).

The process shown in Figure 3, produces 63.5 g/L of succinate per batch at a production rate of $2.54 \text{ g L}^{-1} \text{ hr}^{-1}$ (Zhu et al., 2011) in fermentation. The broth is separated from the cells in a disc-stack centrifuge prior to adding sulfuric acid to precipitate calcium sulfate and making succinic acid (Fujita and Wada, 2011). While this process does produce a stream of non-product, current alternatives prove too energy intensive (Glassner and Datta, 1989). However the gypsum may have use in a variety of industries such as food preservation, concrete, and at sufficient purity biomedical applications (Aguilera and Karel, 1997; Aïtcin, 2000; Coetzee, 1980; Kurzrock and Weuster-Botz, 2010). To conduct this purification process, the initial gypsum is removed with the first crystallizer and disc-stack centrifuge followed by crystallization of succinic acid removal and drying. The crystal solution has inverse solubility relationships between calcium sulfate and succinic acid that constrains the temperatures and order of separation of these two chemicals (Fujita and Wada, 2011).

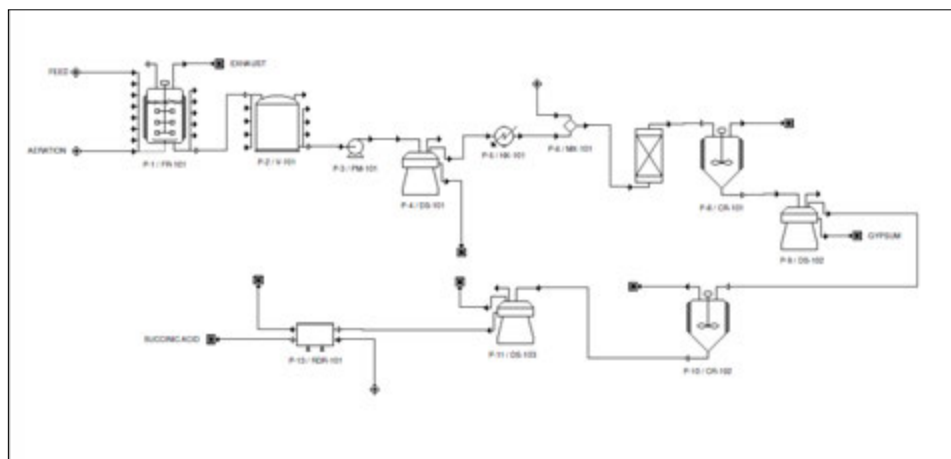


Figure 3: SuperPro process flow diagram for succinic acid

BioPET again uses inputs to mimic Figure 3, and these are shown in Table 12 for succinic acid. Of the key differences, BioPET does not include centrifugation steps for the each crystallizer process that can potentially underestimate the process economics.

Table 12: BioPET Inputs – Succinic Acid

Input Variable	Value	Source
Production Size	15 KTA	
Operating Days	330 days	
Internal Rate of Return	10%	
Plant Operating Life	10 years	
Mass Ratio	1	
Fermentation		
Productivity	2.54 g/L/hr	(Zhu et al., 2011)
Titer	63.5 g/L	(Zhu et al., 2011)
Yield	1.049 g/g	(Zhu et al., 2011)
Separation	N/A	
Catalysis A -> C	N/A	
Primary Purification	Crystallization	
Mass Separation	1 kg _{agent} /kg _{product}	
Purity	0.5	(Fujita and Wada, 2011)
Yield	99%	(Fujita and Wada, 2011)
Secondary Purification	Crystallization	
Energy Separation		(Fujita and Wada, 2011)
Purity	0.5	(Fujita and Wada, 2011)
Yield	99.9%	(Fujita and Wada, 2011)

Model Comparison: Adipic Acid

Adipic acid is a valuable compound that represents a precursor to 6,6-nylon, a petrochemical that is steadily increasing in cost. It has been suggested that bio-production is expected to be cost-neutral with a crude oil price of approximately 40 \$/bbl (Guzman, 2010). While limited data exists on microorganisms synthesizing cis-cis muconic acid, a precursor to adipic acid, fermentation values for productivity and titer were chosen in the range of prior chemicals. Conservative values are 2 g/L/hr and 40 g/L respectively. This titer is near optimistic values for the BREW project and experiences near ethanol-like productivity values (Patel, 2006). Cis-cis muconic acid yield was assumed to be 90% of theoretical maximum on a molar basis (Patel, 2006). From cis-cis muconic acid, hydrogenation has been demonstrated over a platinum catalyst to yield adipic acid (Draths and Frost, 1994). For purification of adipic acid from the catalyst effluent, exploitation of the temperature sensitivity of adipic acid was assumed to be the best method of separation (Musser, 2000). The assumed PFD is shown in Figure 4 and describes everything from fermentation, catalysis, to final purification.

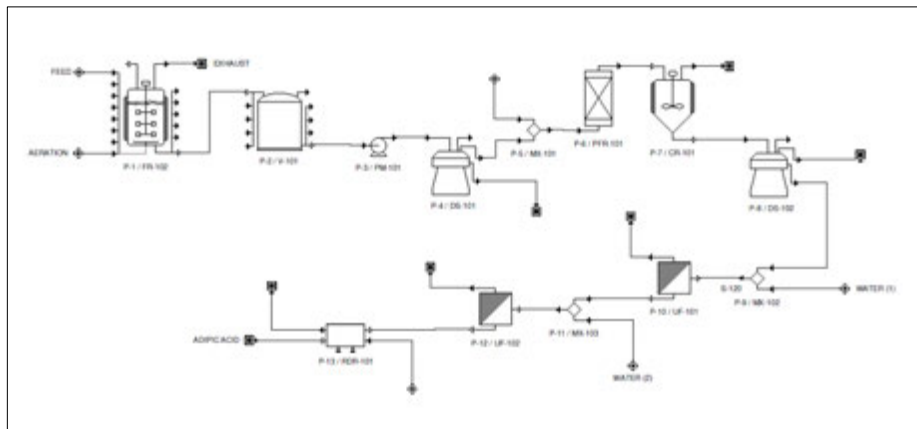


Figure 4: SuperPro process flow diagram for adipic acid

BioPET is able to mimic Figure 4 with the inputs listed in Table 13 and is also able to account for the catalyst that is not directly accounted for in SuperPro without extra

programming. The mass ratio accounts for the increase in, or in other cases a decrease, mass added to the cis-cis muconic acid to convert it to adipic acid. The mass of adipic acid is 1.03 times that of cis-cis muconic acid.

Table 13: BioPET inputs – Adipic Acid

Input Variable	Value	Source
Production Size	15 KTA	
Operating Days	330 days	
Internal Rate of Return	10%	
Plant Operating Life	10 years	
Mass Ratio	1.0283	
Fermentation		
Productivity	2 g/L/hr	
Titer	40 g/L	(Hermann and Patel, 2007)
Yield	.47 g/g	(Patel, 2006)
Separation	N/A	
Catalysis A	Hydrogenation on Platinum	(Draths and Frost, 1994)
Selectivity	99%	(Draths and Frost, 1994)
Conversion	90%	(Draths and Frost, 1994)
Catalyst Life	5-years	
Primary Purification	Crystallization	
Energy Separation		
Purity	0.5	
Yield	99.9%	

Ethanol Comparison

For overall cost, BioPET produced an ethanol production cost estimate within 1% of SuperPro Designer[®] under the same assumptions. The majority of the cost was a result of the feedstock for both models. The most significant differences were in the capital and utilities costs. Nutrients and labor were both within 5% of each other and appeared to produce adequate results.

Table 14: Comparison of SuperPro and BioPET output for ethanol production.

Ethanol	SuperPro		BioPET		Error Between Models (%)
Item	Value (\$/kg)	% of total cost	Value (\$/kg)	% of total cost	
Feedstock	0.309	59%	0.297	57%	3.8%
Capital + Overhead	0.043	8%	0.147	28%	241.9%
Nutrients	0.062	12%	0.035	7%	43.5%
Labor	0.043	8%	0.024	5%	44.1%
Utilities	0.065	12%	0.017	3%	73.8%
Total	0.52	100%	0.52	100%	0.0%

An additional benefit of BioPET was its ability to discretize costs according to the major steps within a process. The breakout of these costs is listed Table 15 and produces feedstock costs closer to 70%; similar to the described cost by Cyweski and Wilke (1978). The next major cost was the result of fermentation; which must account for fermenters, centrifuges, and micronutrients. Separation accounts for the least amount in the production of ethanol, but represents the largest amount of utilities consumed.

Table 15: Ethanol Results: BioPET

Ethanol	Value (\$/kg)	%
Feedstock	0.36	69%
Fermentation Cost	0.13	24%
Separation Cost	0.04	7%
Total	0.52	100%

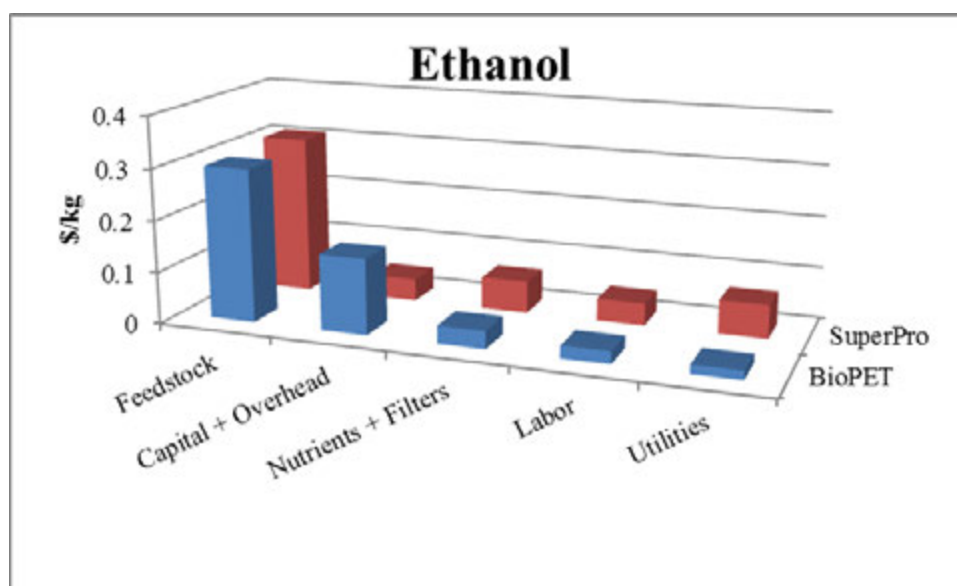


Figure 5: Comparison of the economics between BioPET and SuperPro process models for ethanol

Succinic Acid Comparison

The BioPET analysis of succinic acid did not produce the same cost, but was within 1% of SuperPro estimates. Feedstock estimates were again near identical, with utilities and capital differing by less than 5%. The largest margin of difference was a result of the nutrient and separating agents and the costs of labor. The large labor costs from SuperPro arise out of the multiple separation steps, which are categorized as labor intensive.

Table 16: Succinic Acid

Succinic Acid	SuperPro		BioPET		Error Between Models (%)
Item	Value (\$/kg)	%	Value (\$/kg)	%	
Feedstock	0.141	12%	0.143	12%	1.4%
Capital + Overhead	0.526	43%	0.463	38%	12.0%
Nutrients + Separating Agents	0.169	14%	0.351	28%	107.7%
Labor	0.281	23%	0.206	17%	26.7%
Utilities	0.096	8%	0.061	5%	36.5%
Total	1.21	100%	1.23	100%	1.7%

The BioPET breakout of costs placed minimal emphasis on the feedstock as a result of the large yield of succinic acid on glucose and carbon dioxide. The largest cost is a derivative of separating succinic acid in a two-step crystallization process, which in turn also incorporates the largest number of unit operations. Fermentation accounted for an almost equal percentage of final product cost as in ethanol.

Table 17: Succinic Acid Results: BioPET

Succinic Acid	Value (\$/kg)	%
Feedstock	0.17	14%
Fermentation Cost	0.32	26%
Separation Cost	0.74	60%
Total	1.23	100%

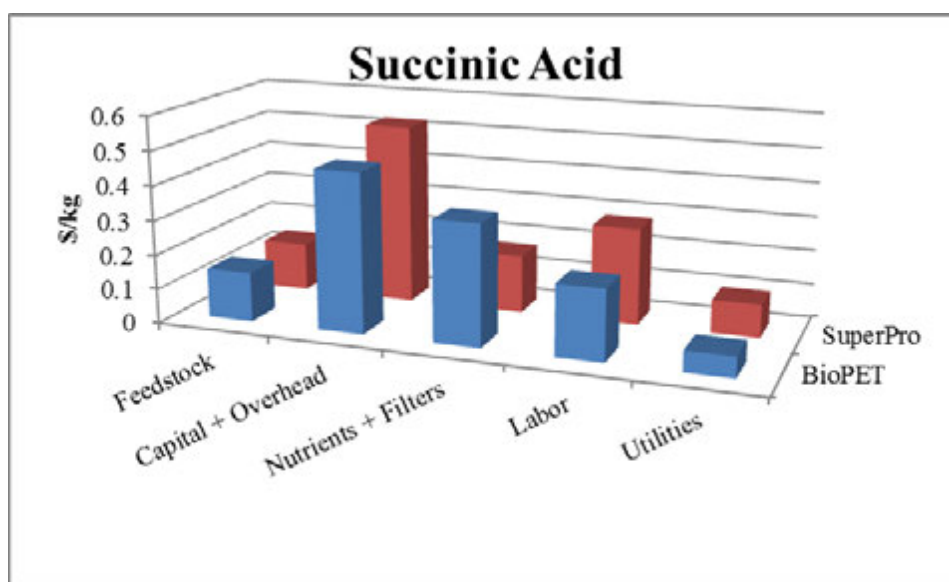


Figure 6: Comparison of the economics between BioPET and SuperPro process models for succinic acid

Adipic Acid Comparison

The comparisons in a more complex process that not only includes fermentation and separation but also catalysis produced the largest amount of variance between SuperPro and

BioPET. Feedstock, nutrients, and utilities all were within 5% of each other. Larger variations occurred in capital and overhead, labor, and catalyst costs. This process also had the largest discrepancy of cost to producing the main product. The costs are similar though and represent similar performance at an early-stage analysis.

Table 18: Adipic Acid

Adipic Acid	SuperPro		BioPET		Error Between Models (%)
Item	Value (\$/kg)	%	Value (\$/kg)	%	
Feedstock	0.379	28%	0.349	24%	7.9%
Capital + Overhead	0.424	24%	0.501	35%	18.2%
Nutrients	0.136	11%	0.088	6%	35.2%
Labor	0.401	34%	0.155	11%	61.3%
Utilities	0.045	3%	0.075	5%	66.7%
Catalyst	-	-	0.276	19%	
Total	1.39	100%	1.44	100%	3.6%

BioPET described a cost of adipic acid to be nearly evenly split between feedstock, fermentation, separation, and catalyst cost. Fermentation produced a 26% of final product cost again. The separation represents a decreased portion of the cost with fewer unit operations selected within BioPET. Catalyst costs, while the most uncertain due to the minimal input to BioPET, represented a significant portion of the final cost.

Table 19: Adipic Acid Results: BioPET

Adipic Acid	Value (\$/kg)	%
Feedstock	0.42	29%
Fermentation Cost	0.38	26%
Separation Cost	0.24	17%
Catalysis Cost	0.40	28%
Total	1.44	100%

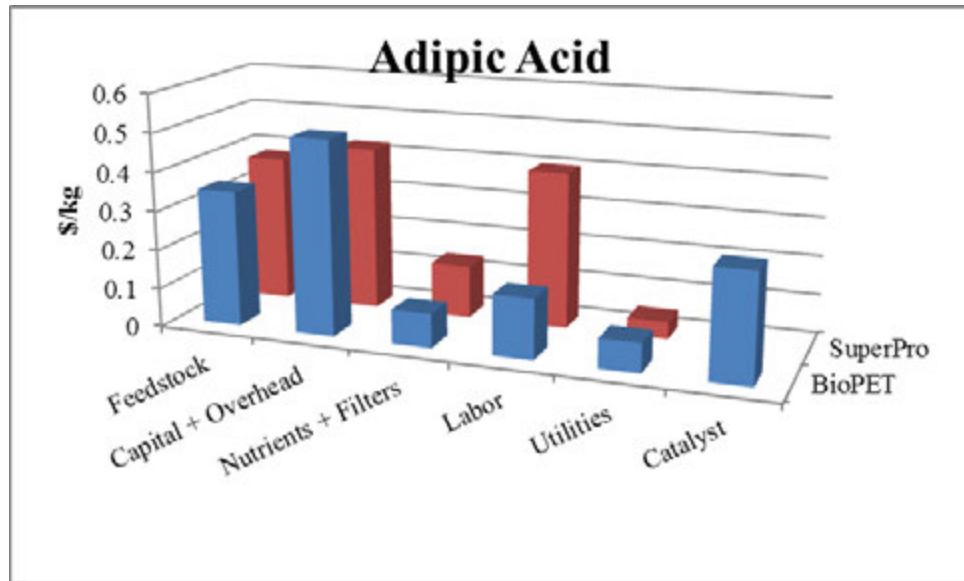


Figure 7: Comparison of the economics between BioPET and SuperPro process models for adipic acid

Discussion

BioPET vs. SuperPro

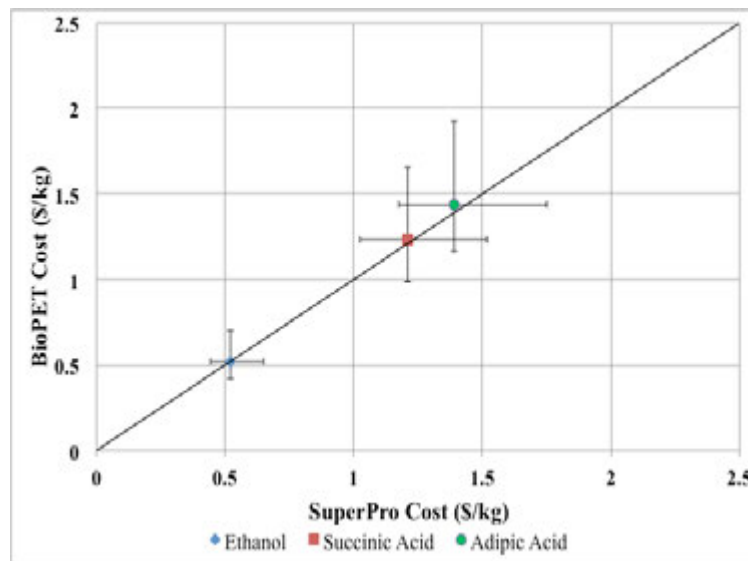


Figure 8: Regression of predicted selling price for ethanol, succinic acid, and adipic acid for BioPET and SuperPro Designer® modeling tools (SuperPro error bars, +25/-15%; BioPET error bars, +35/-20%)

Examining the final production cost of each product and the expected ranges of uncertainty in product cost, Figure 8 illustrates how preliminary cost estimation can afford a reasonable prediction of cost at early-stages in process development. These error bars are often within the realm of expected uncertainty as processes begin their progress towards commercialization and will gradually narrow as more detailed information is acquired.

The largest source of error tended to be in operating expenses as opposed to fixed costs which can be expected as operating expenditures tend to have a larger impact on final cost due to the annual occurrence versus the one-time expenditure that is amortized. Another portion that is difficult to account for is the rapid automation that can be seen by much of the petrochemical industry that would result in decreased labor cost with minimal increases in fixed costs. However, to account for automation would require industrial survey and may potentially provide labor, albeit maybe more accurate, but not representative of any commercial modeling program available.

The major benefit of a tool such as BioPET is the ability to generalize bioprocessing for future bio-based products and biorefinery innovations with instantaneous feedback. This model may only consider one product stream and no feedstock conversion, but may see future adaptations to address this. A multiple-product biorefinery is not a business model that has been adopted yet, but could potentially lead to more economically feasible production of biofuels with value-added products such as petrochemical drop-in chemicals.

Conducting a sensitivity analysis on the production size of the facility modeled in BioPET as well as the feedstock cost produced fairly linear relationships due to the linear nature of the majority of the model. The sensitivity of production size shown in Figure 9, appears in line

with published data (Gallagher et al., 2005) with an overall plant exponent for ethanol of 0.83. The other processes appear to reach a minimum and then experience a cost that fluctuates more as a result of the construction of BioPET rather than scale. This may describe a process that is more attached to equipment and overhead than the extremely feedstock dependent process of ethanol. Fitting trendlines through the data in Figure 9 using a power law produced poor fitting equations ($R^2 < 0.7$) for the non-ethanol processes whereas ethanol approached an R^2 of 0.94.

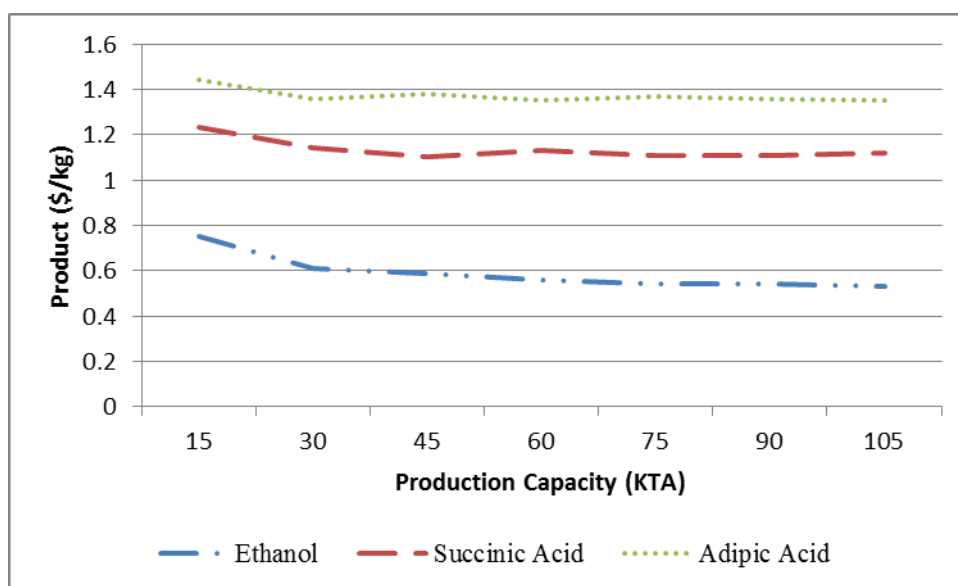


Figure 9: Sensitivity of production capacity to selling price for ethanol, succinic acid, and adipic acid using the BioPET tool

Figure 10 depicts the sensitivity to feedstock cost from BioPET with succinic acid being the least sensitive. This is a result of a higher yield of the product on glucose than the other processes and appears in accord with predictions about costs (Guzman, 2010). This also shows the power of a preliminary estimation of product cost to provide a range of costs and ballpark a sensitivity to key parameters such as feedstock cost.

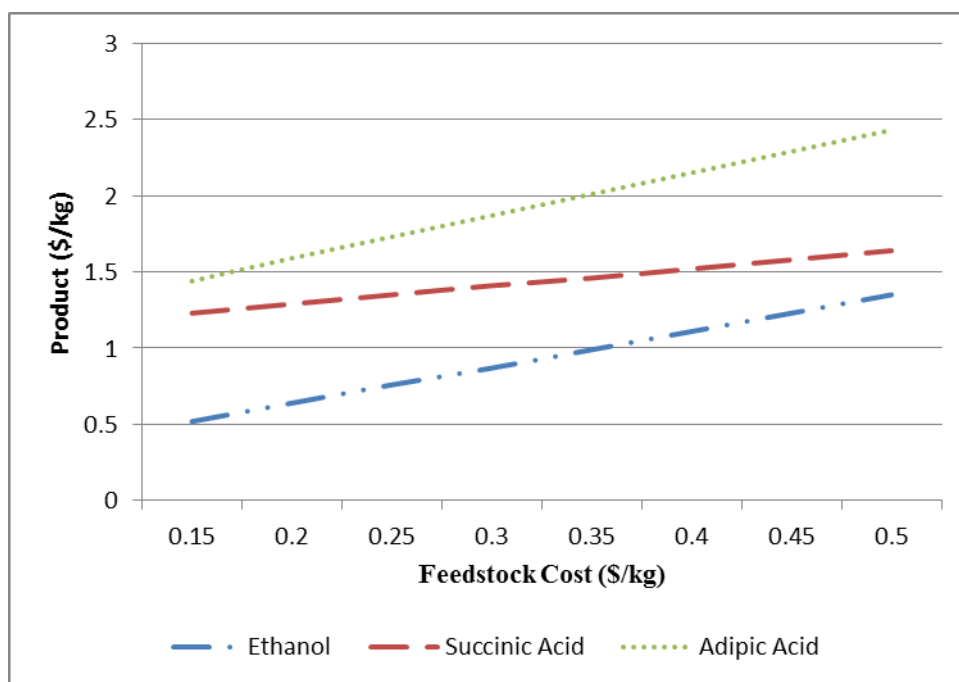


Figure 10: Sensitivity of feedstock cost for ethanol, succinic acid, and adipic acid using the BioPET tool

BioPET vs. BREW Project

The accuracy of the BREW project can provide a certain level of validation of the accuracy of BioPET. While exact results should not be expected as is seen in many other TEA's, a general trend should at least be visible considering the likeness of the studies. The greatest difference between the cost estimated by the BREW project and the cost estimated by BioPET was the ethanol process. This is most likely due to the lack of extra processing of co-products that often accompanies ethanol processing which is supported by the consistent difference between the two costs across all ranges of fermentable sugar cost. Succinic acid has near identical estimations to the BREW project, while adipic acid approached BREW estimates at higher levels of feedstock cost. BioPET predicts adipic acid may be more sensitive to feedstock costs than that of the BREW project analyses.

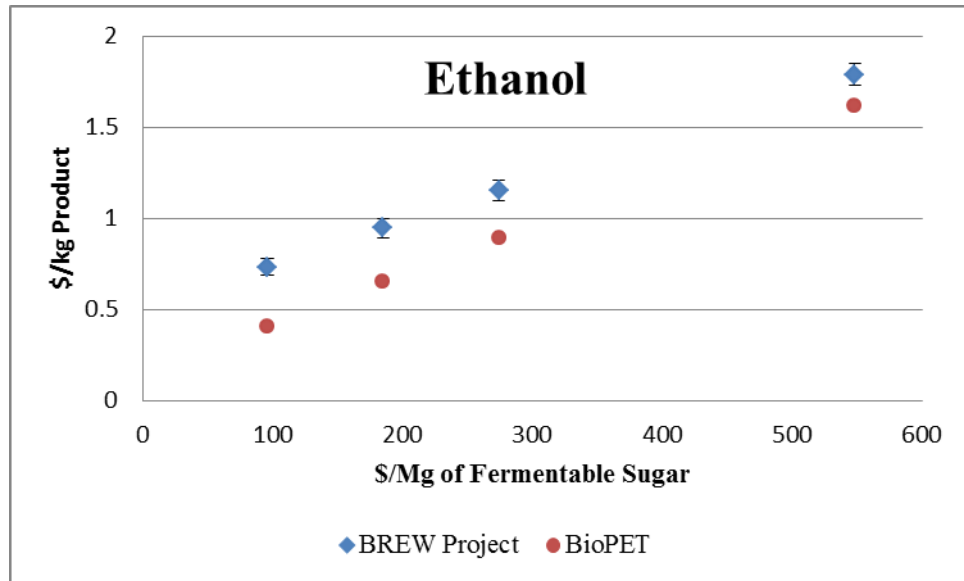


Figure 11: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for ethanol.

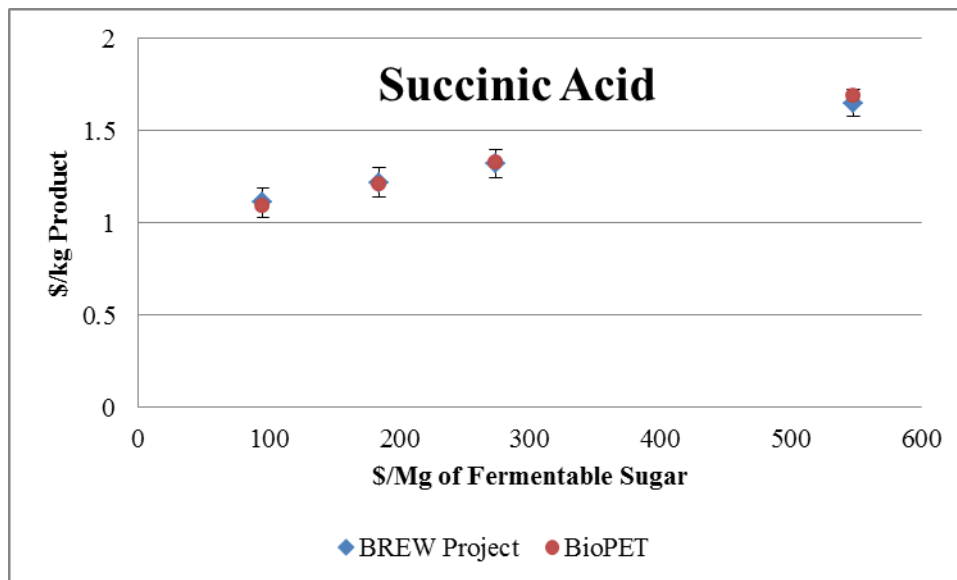


Figure 12: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for succinic acid.

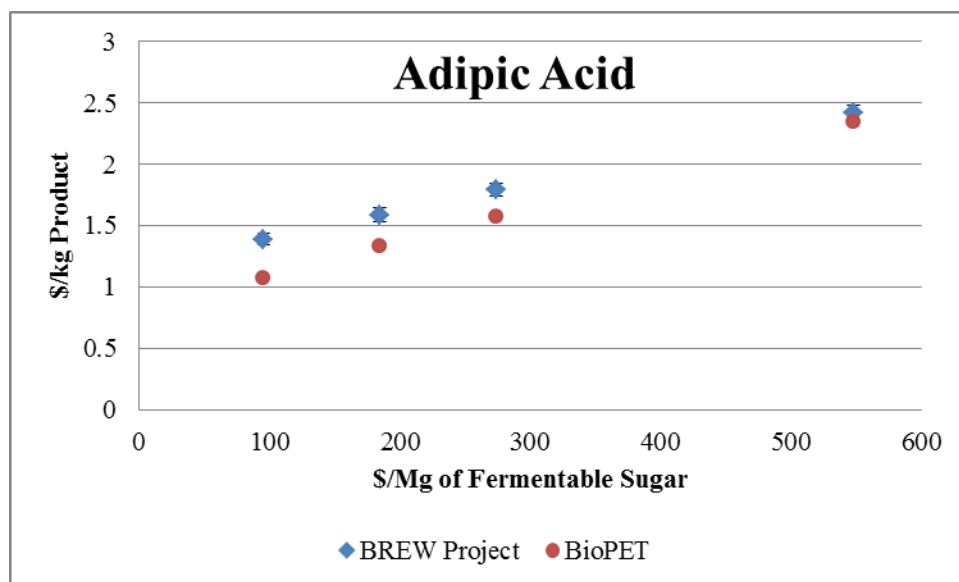


Figure 13: Sensitivity of BioPET and BREW project models to different purchase prices of fermentable sugars for adipic acid.

Conclusion

A new tool for preliminary cost estimation has been developed and tested against modeling software and published results of previous cost-estimations for these processes. The results displayed a tool capable of predicting feedstock and capital cost near identical to programs such as SuperPro Designer[®]. While incapable of replacing a program such as SuperPro Designer[®] due to the level of detail that a modeling program as such can produce, the value of preliminary models for early-stage process development has been shown. BioPET has also been shown to produce results within the range of previous studies helping further validate this technique. To improve further preliminary estimations of bio-based products, it would be a valuable tool to add in items such as a feedstock decisions and an integrated life-cycle assessment to promote understanding of economic and life-cycle tradeoffs of a new chemical processes.

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CHAPTER 4: TECHNOECONOMIC EVALUTION OF BIO-BASED STYRENE PRODUCTION FROM *ESCHERICHIA COLI*

Introduction

Biorenewable fuels and chemicals have received a great deal of attention in the recent years due to political and economic concerns over depleting petroleum supplies (Kazi et al., 2011; Rogner, 2012). Although the traditional focus has largely been on the development of fuels from renewable sources, such as ethanol or biodiesel, a new surge of research has focused on the production of chemicals as “value-added” products from biomass (Werpy and Petersen, 2004). Many methods for producing bulk chemicals from biomass have been proposed and/or successfully implemented, including thermochemical, biochemical, and catalytic approaches, as well as hybrids of these methods (Brown et al., 2012; Kazi et al., 2011; Nikolau et al., 2008; Xie et al., 2006). Bulk chemicals typically have a slightly to greatly higher value per unit mass than do fuels, and serve markets that are significantly smaller on the basis of total demand for carbon. This is exemplified by the observation that the bulk chemicals market is approximately the same economic value as the fuels market, despite using only 1/20th as much carbon (Nikolau et al., 2008).

Chemical production via biochemical routes has grown from the historical success of ethanol fermentation, while benefitting from ongoing improvements in the fields of synthetic biology and metabolic engineering. Ever-expanding genetic toolkits and novel predictive tools, have led to enhanced fermentation kinetics, elevated titers and yields, improved tolerance to environmental stresses and product-toxicities, and novel metabolic pathways for the production of non-natural compounds (Nielsen, 2001). Several products have been successfully

commercialized with the aid of these new technologies; including polylactide (NatureWorks™), 1, 3 propanediol (DuPont), and succinic acid (BioAmber). Many other bio-based chemicals in the development pipeline may also have the potential to one-day reach commercial success; however, not before significant improvements can be made with respect to critical biological parameters and cost-effective scale-up methods. Technoeconomic assessments (TEA's) can help to bridge the gap between research and commercialization by illuminating bottlenecks and opportunities in a bioprocessing scheme (Hermann and Patel, 2007; Kazi et al., 2010; Kazi et al., 2011). A challenge for TEA is that these studies usually require significantly more information than is readily available in the early-stages of process development. Such process uncertainties can cause errors in the analysis that the reader and the author may not be aware of (Bunger, 2012). To overcome these opaque and complex models, simplified models can instead be developed to generate a greater level of transparency of the various assumptions and inherent uncertainty in the process evaluation with minor sacrifices in the precision of the estimate. In cases relevant to early-stages of development, a simpler model can perform nearly as well as the commercially available tools such as SuperPro Designer® (Claypool, 2013).

Although it remains in the early stages of development, styrene has recently emerged as a commercially viable bio-derived chemical candidate with great potential (McKenna and Nielsen, 2011). With an annual consumption greater than 5.8 million metric tons in the United States alone, styrene is an important monomer and platform chemical used across many different industries, and is produced primarily from petroleum-derived ethylene and benzene (Chen, 2000). With current styrene prices in the range of 1.74 – 1.83 USD kg⁻¹, and future price hikes expected, some authors suggest bio-based styrene may become an economic alternative (Balboa, 2013). While the current state of bio-based styrene is far from commercial-scale production,

investigation into the economic performance of this new bioproduction pathway is of great interest to the research community and industry alike.

Through *de novo* pathway design, the non-natural styrene biosynthesis pathway was recently engineered using the bacterium *Escherichia coli* as the biocatalyst platform (McKenna and Nielsen, 2011). By extension of the endogenous L-phenylalanine pathway, styrene is produced via *trans*-cinnamate with the aid of two heterologous enzymes (McKenna and Nielsen, 2011). As the maximum theoretical yield of L-phenylalanine on glucose is 0.55 g/g, styrene could be produced at a maximum theoretical yield ranging from 0.26 to 0.346 g/g (Báez-Viveros et al., 2004; McKenna and Nielsen, 2011; Nielsen, 2013). A significant problem currently hindering high-titer production of bio-styrene arises due to its significant toxicity against *E. coli* (McKenna and Nielsen, 2011). The toxicity limit against *E. coli* has been predicted to be ~300 mg/L, and with maximum titers by first-generation strains already approaching this concentration, toxicity looms as a critical limiting factor in styrene biosynthesis. However, since the solubility of styrene in water is a mere 320 mg/L at operating temperatures, if the *E. coli* can be engineered to withstand only slightly higher styrene concentrations, spontaneous phase separation would simultaneously ensure toxic concentrations would no longer limit bio-styrene production while greatly facilitating downstream - or even in situ - product recovery. Furthermore, as water is highly insoluble in styrene the resultant product would consist of extremely high purity styrene (>99.8%), suitable for most polymer standards (Chen, 2000).

In this work we have applied the Biorenewables Process Evaluation Tool (BioPET), which is a spreadsheet-based tool for early-stage evaluation of biorenewable processes, to examine the potential of bio-styrene production, and to illustrate key process bottlenecks (Claypool, 2013). Combining the available knowledge of the physical properties of styrene, as

well as factors relevant to and influencing styrene biosynthesis, BioPET can facilitate an evaluation of the commercial-scale economics of such a venture. This paper will investigate what benchmarks will need to be reached for commercialization of biorenewable styrene and the likely minimum estimated selling price (MESP) anticipated through such a route.

Methods

Styrene bioproduction will incorporate both fermentation and separation processes. The method of separation at commercial-scale is likely to exploit the mutual insolubilities of styrene and water, much like between fatty acid esters and glycerin in biodiesel production (Marchetti et al., 2008). To implement this economic analysis, the framework of BioPET was chosen and modified to include a decanter separation technique downstream of fermentation. Using published rules of thumb and economic equations as previously presented in BioPET, a decanter was sized and cost estimations provided (Claypool, 2013; Woods, 2007).

The decanter was assumed to operate at 32°C and to have two phases represented by styrene and water. The two phases assume all properties of water and styrene respectively and a bubble diameter of styrene in the continuous water phase of 150 µm. As the properties of these two components led to a settling velocity greater than 4 mm/s, a terminal settling velocity of 4 mm/s was chosen. Due to expected low flow rates over the range of operation and the large settling velocity, a vertical column decanter was assumed. The necessary area of interface between the two phases was calculated using Equation 1.

$$A_i = \frac{u_d}{v_c} \quad (1)$$

Where A_i = the area of interface, m^2 ,

u_d = the terminal settling velocity, m/s,

V_c = the volumetric flow rate of the continuous phase, m³/s.

This area was assumed to be the cross-sectional area of the column with an L/D of 5. The cost calculations were then as follows using values from a previous-described decanter with a continuous phase flow rate of 12 L/s at an original adjusted cost for stainless 304 of 285,000 USD (Woods, 2007). The exponential relationship for decanters follows an exponent of 0.84 and follows Equation 2 (Peters et al., 2003).

$$C_n = \frac{S_n}{S_o} \times C_o^n \quad (2)$$

Where:

C_n = new cost for newly sized piece of equipment

S_n = new size of equipment

S_o = size of equipment where previous cost data exists

C_o = cost of equipment where previous data exists

n = empirically-derived cost exponent

This equation uses the previous knowledge of the decanter cost and applies it over a range of continuous phase flow rates between 1.2 and 40 L/s (Woods, 2007).

A styrene bioproduction process was designed using the expected values in and examined over the entire range, worst-case to best-case, using a Monte Carlo approach with 2000 simulations.

Table 1: Base-Case (expected value) and range of parameter values for Monte-Carlo analysis of the economics of styrene bioproduction.

Parameters	Expected Value	Worst-Case for Value	Best Case for Value
Annual Production (kilotonnes/yr)	45	22.5	67.5
Operating Days (days)	345	327.75	362.25
Internal Rate of Return (%)	15	21	9
Plant Operating Life (yrs)	7	3.5	10.5
Lang Factor (dimensionless)	5	6.5	3.5
Productivity (g/L/hr)	2	1.4	2.6
Titer (g/L)	50	35	65
Yield (g/g)	0.25	0.225	0.275
Density of Product (g/L)	0.888	0.879	0.897
Product Purity (wt.%)	0.999	0.998	0.999
Product Solubility (wt.%)	0.4×10^{-3}	0.52×10^{-3}	0.28×10^{-3}
Glucose (\$/kg)	0.3	0.45	0.15
Corn Steep Liquor (\$/kg)	0.2	0.3	0.1

Results and Discussion

Using the base-case assumptions (leftmost numerical values in Table 1), BioPET estimated an MESP of 1.82 USD kg⁻¹ for 99.9% pure styrene monomer. It is predicted that the process will employ two fermenters of approximately 2040 m³ each operating for 276 batches per year. Due to the estimated ease of separation, it is expected that a single decanter should be adequate for the size and titer values estimated. No alternative products are considered to be produced or interfere with product purity (McKenna and Nielsen, 2011).

Table 2: Capital Expenditures

Section	Capital Costs (in Million USD)
Fermentation	14.7
-Fermenters	10.1
-Seed Fermenters	3.7
-Holding Tanks	0.9
Separation	6.8
-Centrifuges	4.2
-Decanters	2.6

The total capital investment for all installed equipment for the expected values was 21.5 million USD with the worst-case values estimating a cost of 19.3 million USD, because the worst-case scenario assumed annual production of 50% of base-case. The capital requirements *per unit annual production* are nearly twice as large for the worst-case scenario. The largest capital expenditure arose out of fermentation with an installed cost of 14.7 million USD as seen in Table 2. These expenditures are under the estimated Lang Factor of 5, a method that aggregates all installation factors into a single multiplier, which is in accordance with suggested liquid-liquid processing systems, but may be closer to 3 as is estimated for corn-grain ethanol (National Renewable Energy et al., 2000; Peters et al., 2003) – we therefore consider our capital cost estimate to be fairly conservative. Lack of prior information to further guide the estimates is not available and leads to a large range of uncertainty.

Table 3: Annual Expenses

Section	Annual Expenses (in Million USD)
Feedstock	54.0
Fermentation	7.3
-Amortized Capital	3.5
-Utilities and Nutrients (non-feedstock)	3.8
Separation	2.4
-Amortized Capital	2.0
-Utilities	0.3
Plant Expenses	15.9
-Plant Overhead	7.4
-General Expenses	6.8
-Maintenance, Patents, Operating Supplies	1.7
Labor	2.2

Feedstock was a dominating factor in the operation of the styrene plant accounting for 67%, or 54 Million USD, of the total annual expenditures, as seen in Table 3. In line with the comparison to biodiesel, simple processing systems become heavily dependent on feedstock and can be estimated at upwards of 90% of the annual expenses (Marchetti et al., 2008). These estimates from BioPET for bio-based styrene also result in approximately similar breakdown of costs as a percentage, as the estimate of ethanol from BioPET (Claypool, 2013). The cost of feedstock (assumed to be pure glucose) corresponds to approximately 6.00 USD bu⁻¹, or 240 USD Mg⁻¹. Although any glucose feedstock could be chosen, if corn was the resulting choice the required number of bushels would be approximately 60,000 acres of corn or <1% of the harvested land in Iowa for corn grain in 2011 (Department of Agriculture, 2011). If these statistics hold true for future years, this would correspond to a styrene plant in Iowa being able to acquire all necessary feedstock within an 8.8-mile radius.

As uncertainty exists both in the BioPET and the assumed values, both a sensitivity analysis and Monte Carlo simulation were conducted to evaluate the validity and sensitive points of error. The sensitivity analysis was conducted by adjusting parameters individually by $\pm 1\%$, and measuring a percent change in the output of the MESP. The top five most sensitive parameters were identified and displayed in Figure 1. The usable fraction of fermenter, assumed to be used at 80% of total volume, was the most sensitive parameter. The 80% value was assumed to leave head room for foaming and aeration issues as done by other authors (Cysewski and Wilke, 1978). Yield of product on substrate was another key parameter with significant uncertainty stemming from the uncertainty in the yield of L-phenylalanine (Nielsen, 2013).

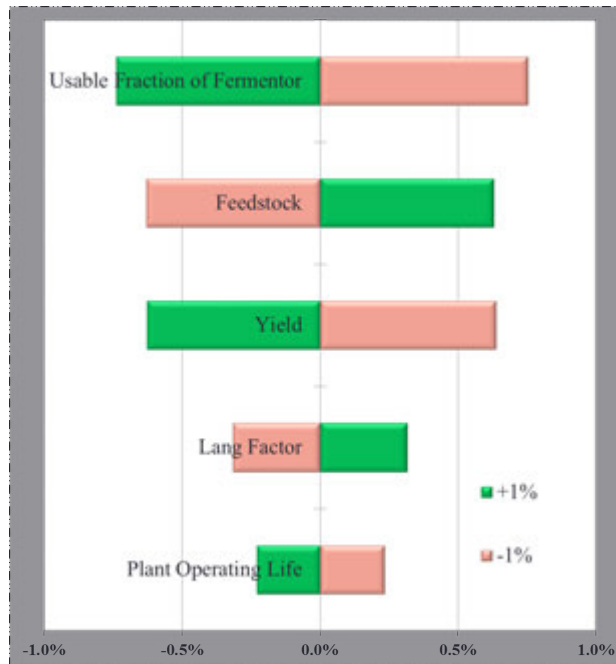


Figure 1: Top five sensitivity coefficients using a $\pm 1\%$ change in BioPET model under expected values for bioproduction of styrene (x-axis represents measured change in MESP).

The Monte Carlo simulation produced an estimated MESP of 1.82 USD kg^{-1} product with a standard deviation of 0.44 USD kg^{-1} . As shown in Figure 1, the yield of product on substrate and cost of feedstock are major driving factors for the large variation in the MESP.

Overall, the economic analysis of bioproduction of styrene expressed potential commercial feasibility at modest fermentation productivities, titers, and yields. Smaller bio-based chemical plants, such as succinic acid, are being built at annual production rates of 34 Mg per year, but larger plants will most likely be constructed because of the benefits of economies of scale (BioAmber, 2012; Haldi and Whitcomb, 1967). It seems likely that production values out of fermentation can likely exceed predicted values due to the phase separation limiting product inhibition, which can be seen in other fermentations where no phase separation occurs (Levenspiel, 1980). However toxicity is currently the limiting case to production and must be overcome to achieve any competitive commercial values.

Toxicity has presented issues historically and one method of keen interest to researchers is to use *in-situ* extraction, or *in-situ product removal* (ISPR), that extracts the toxic product of interest into a second phase, typically a biocompatible solvent, in order to limit the effect of the toxic product (Brennan et al., 2012). While the economics of ISPR might become practical when the product of interest is high-value, the cost of biocompatible ISPR agents and their respective reduction in fermentation volumes per purchased volume, it does not seem a prospective path for bio-based styrene. Alternative methods of overcoming the toxicity must be sought (Jarboe et al., 2011).

Another key risk is feedstock purchase price. Over the past decade, corn prices have ranged 1.75 to nearly 7 USD bu⁻¹ (four-fold) while oil prices have ranged from approximately 25 – 125 USD bbl⁻¹ (five-fold). This may be a weak spot in the future of styrene in that the economics are sensitive to volatile feedstock prices, but with improvements in the decomposition of lignocellulosic feedstocks, these new feedstocks may prove cost competitive for fermentable sugars (Rezaei et al., 2011). Projected costs for these lignocellulosic feedstocks have even been

estimated as being three times less expensive than corn starch at 2.50 USD bu⁻¹ from corn grain (Lynd et al., 1999). It seems probable that between corn grains historical price and estimated prices for lignocellulosic sugars, the estimated cost to produce bio-based styrene has a potential future with a variety of feedstocks.

While the major costs are associated with feedstock and yields, alternative driving factors are capital costs. Bio-based styrene represents a fairly simple process design that mimics corn-grain ethanol in that the general process consists of fermentation and a single separation unit operation. Another factor that may also mimic ethanol is the Lang Factor of which will reduce the estimates to produce bio-based styrene (National Renewable Energy et al., 2000). The push to create new bio-based products can have a major impact on its own industry by driving the capital costs of fermenters down via an increased available supply, although the demand may outweigh this benefit.

Conclusion

A 45 kilotonne per annum bio-based styrene plant is estimated to produce 99.9% pure styrene monomer at a MESP of 1.82 USD kg⁻¹. This price is competitive with current styrene monomer prices in today's market (Balboa, 2013). Considering uncertainties of details around final construction costs and key fermentation parameters, the estimate for bio-based styrene is 1.82 ± 0.44 USD kg⁻¹. This matches current market values for styrene and presents a great opportunity for investment while still possessing a great amount of risk. While an overarching amount of uncertainty in product yield can limit the competitiveness of the future selling price of commercial-scale bio-based styrene, initial toxicity presents a barricade to achieving the necessary production values. Future research should be targeted towards addressing or overcoming this limitation.

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CHAPTER 5: GENERAL CONCLUSIONS

The field of industrial biotechnology is growing rapidly. Yet despite the gains in understanding subjects from enzyme design to metabolic engineering, early-stage technoeconomic evaluation remains an area that has been poorly explored. In this work, a robust and easy-to-use spreadsheet-based TEA for biorenewable chemicals was developed and validated. Chapter 2 examines a preliminary spreadsheet-based approach for analyzing the economics of a combined fermentative-catalytic route to producing sorbic acid. This spreadsheet-based analysis demonstrated the feasibility of utilizing a coarse technoeconomic approach for evaluating processes within this new industry. The model helped illustrate the importance of overall yield because of the cascading effect that low yield has on the required size of all upstream processes. With the development of metabolic engineering, it becomes increasingly likely that more of the chemical synthesis can be done *in vivo*, and trade-offs between capital costs, productivity, and yield must be considered carefully. The cost predicted out of the model suggested a plausible outcome in the long-term scenario for sorbic acid. The plausible outcome for such a scenario suggests that future work should continue on such a project.

Influenced by Chapter 2, Chapter 3 describes the development of a more general model for evaluating bio-based chemical production. Based on the Chapter 2 experience, a common framework was employed, involving: Fermentation, Separation, Catalysis, and Purification. These key processes defined the structure of the resulting model, which was termed BioPET (Biorenewables Process Evaluation Tool). The level of detail available at early stages in development dictated that BioPET be operable with relatively minimal amounts of process detail. While tools are only as good as the operator and the information they are provided,

BioPET was built for ease of use and validated against SuperPro Designer[®] to reflect how given the same information, similar outputs could be generated. BioPET was also analyzed against other literature values in order to measure the accuracy at which it could predict values. The analysis against literature and alternative software provided feedback on both the precision and accuracy of the tool at hand and demonstrated good correlation for both. This tool also demonstrated the potential variability in the cost distribution of bio-based chemicals.

To demonstrate the potential of BioPET for predicting costs on a novel bioprocess, Chapter 4 focused on a chemical that has direct market potential as an intermediate molecule, and further developed BioPET by adding sensitivity analysis and a Monte Carlo simulation into the model. BioPET also required the addition of decantation, a common unit operation for separating immiscible liquids. The addition of the sensitivity analysis and Monte Carlo simulation required the addition of VBA code to generate this level of analysis, but did so without sacrificing the benefit of instantaneous feedback to the user. The economic promise of the bioproduction of styrene was demonstrated through the analysis, as was the significant ($\pm 25\%$) uncertainty associated with the estimate.

Overall a platform has been developed for evaluating bio-based chemicals from which can be adapted to provide early-stage insight into future endeavors. The platform, BioPET, provides rapid feedback, sensitivity and Monte Carlo analysis for the biorenewables industry that may help guide and inform future decisions.

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